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Magnesia Scrubbing Applied to a Coal-Fired Power Plant

Chemico Air Pollution Control Co, New York

Prepared for

Industrial Environmental Research Lab, Research Triangle Park, N C

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MAGNESIA SCRUBBING APPLIED TO A COAL-FIRED POWER PLANT

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Research and Development
Program Report

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MAGNESIA SCRUBBING
APPLIED TO
A COAL-FIRED POWER PLANT

by

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ABSTRACT

A full-size demonstration of the magnesia wet-scrubbing system for Flue Gas Desulfurization (FGD) was conducted on a coal fired utility boiler. The system was designed to desulfurize one-half the flue gas from a 190 MW rated capacity generating unit firing 3.5% sulfur coal. The FGD installation was equipped with a first-stage wet scrubber for particle emissions control followed by the magnesia unit.

Twenty-eight hundred hours of operation were logged at the generating station, and the FGD system's ability to remove 90% of the inlet SO_2 was shown. The particle control capability of the unit was also demonstrated by reducing particle emissions to less than 0.01 gr / SCF when the unit was operated in series with an electrostatic precipitator. In a test program using only the wet scrubbing unit for particle emissions control a collection efficiency of 99.6% was achieved.

Magnesia was regenerated and recycled successfully. The SO_2 produced during regeneration was used to manufacture commercial grade sulfuric acid which was marketed. Correlations were developed to determine SO_2 removal for varying boiler loads and fuel sulfur content, and to control regeneration of acceptable alkali. Several other studies of the process technology and process chemistry were undertaken as part of the work.

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1.0 EXECUTIVE SUMMARY

The*Chemico-Basic magnesia slurry SO_2 recovery process for control of sulfur oxides emissions from fossil fuel fired utility boilers has been demonstrated on an oil fired boiler (The Magnesia Scrubbing Process As Applied To An Oil Fired Boiler, EPA 600/2-75-057). This process, with some modifications developed during early operations in New England, was next installed on a coal fired boiler at Potomac Electric Power Company's Dickerson Generating Station as a system to control both particle and sulfur dioxide emissions from the plant. The regeneration and SO_2 recovery steps of the process continued to be carried out at the Environmental Protection Agency's calcination facility at Essex Chemicals' Rumford Acid Plant in Rhode Island, which had originally been constructed for the New England SO_2 Control Project.

Major distinctions between these two FGD units are divided into process design variation and process chemistry differences. In the former category the principal variations between the design of the previously reported oil fired application and this coal fired boiler system were:

A) The addition of a first stage venturi scrubber for particle control only (no alkali is added to this stage). Flue gas exiting this first stage is lowered in temperature to near the adiabatic saturation point through the normal transport mechanisms.

B) Magnesia alkali added to the second stage for SO_2 control is slurried with centrate returned from the centrifuge instead of process water. Since gas entering the second stage is saturated with water vapor little or no evaporation takes

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*Wherever the term "Chemico" is referred to in this report, the term "Chemico Air Pollution Control Company" shall be substituted therefore.

place; thus it is possible to eliminate fresh water inputs to this stage.

C) The use of a co-current, rather than a counter current dryer to dry the centrifuge cake. This design eliminates at least one problem encountered with the countercurrent design used in oil fired application. Dust exiting with the dryer off-gas can be separated and fed directly to the dryer product discharge without adding the pneumatic conveying equipment necessary with countercurrent units.

D) The use of "wet" fan after the two-stage scrubber unit to provide the induced draft for the venturies instead of a hot, forced draft type. An important element of the program was the taking of flue gas containing coal fly ash directly from the boiler, use of the "wet" fan eliminated the erosion problem associated with dry fans handling dirty flue gas and took advantage of the lowered temperature and volume handled after the scrubbers.

The major difference in process chemistry between the two plants was the production of the magnesium sulfite hexahydrate in this coal fired boiler application rather than the magnesium sulfite trihydrate produced during operations at the oil fired generating plant. Processing of the hexahydrate proved similar in many cases, to the experiences encountered when processing the trihydrate.

Another process chemistry difference was the evidence that magnesium oxysulfates were formed, sometimes in large quantities, during the slaking of MgO with centrate. This resulted in a subsequent reduction in SO_2 removal efficiency.

Because of an overlap with the operations of the oil fired station the regeneration plant was not available to the Dickerson facility when the FGD system was ready for its inaugural operation. The lack of a regeneration facility plus problems discovered during the start up period limited the initial PEPCO operations to less than 1000 hours in the nine months period between September 1973 and June 1974.

On July 1, 1974 when the Rumford facility became available, an operating and development program was initiated at PEPCO's Dickerson Station. In the next six months another 1790 hours of operation were achieved despite a number of new problems. All major objectives were completed during the period:

1) Operations were initiated using regenerated MgO.
2) Particle removal efficiency of 99%+ was verified.
3) SO_2 removal efficiency was measured, improved and verified at 90%.
4) Operating conditions for continuous running were established, with emphasis on MgO slurry mixing, control and makeup.
5) MgO loss points were discovered and corrected.
6) An operations evaluation program was established.
7) Material balance measurements were made.
8) The SO_2 removal correlation developed in the previous study was improved to include the effect of regenerated MgO properties.
9) Preliminary correlations were developed to define operating variable influence on dryer and centrifuge system performance.

10) Laboratory investigations of fundamental system properties were continued.

Some important objectives were only partially attained because of a number of shut-downs in the latter period which limited the time available for operation. Thus, a long duration run was not realized. The longest uninterrupted runs were a twelve day continuous period in October and another of eleven days in December. Despite these limited operations two recycles of the magnesium oxide inventory were achieved during the program.

While a number of causes contributed to the interruptions the major "shut-down" problems was deterioration of the piping in the first and second stage recycle loops of the FGD plant. In the second stage extensive corrosion had occurred in the unprotected steel pipe at some time during the preliminary operations. In the first stage recycle pipe, which was rubber lined, a number of failures occurred which were attributed to the supply of off-specification pipe and fittings during construction.

After a piping replacement program and other repairs to the FGD plant had been completed in preparation for a demonstration run, the project was terminated when a routine overhaul of the power plant's generator system in January 1975 revealed a defect in the turbine case. The resultant five month outage required for repair of the turbine extended past the close-out date of the E.P.A. contract for access to the acid plant and operation of the regeneration facility, thus forcing the end of the project.

Despite the limited operations and lack of extended operating data a number of conclusions concerning the operability at the plant can be gathered from the information that was obtained.

1.1 OPERATIONAL EFFECT OF DESIGN VARIATION FROM THE OIL FIRED CASE

A) Particle removal efficiency for both the case in which gas was first cleaned in an electrostatic precipitator and the case where dirty flue gas was taken directly from the boiler exceeded process guarantees. With the electrostatic precipitator in the circuit an average outlet loading of 0.002 gr/ACF was attained with the unit handling its full rated capacity. When treating dirty flue gas an average removal efficiency of 99.6% was obtained at full load operation. The outlet loading from the scrubber system for this case was 0.011 gr/ACF.

B) Slurrying magnesia feed with centrate posed some problems. It appears that part of the MgO feed is complexed as a magnesium oxysulfate. This resulted in the formation of lumps and gels which plugged feed lines in the early operations. This problem was solved by:

1) Providing an intensive mixer between the MgO feeder and slurry tank.
2) Slurrying the MgO feed with the entire recycle centrate stream to reduce concentration.
3) Adding a steam sparger to the slurry tank to heat the slurry to 180°F.

These fixes diminished the problems to a level that allowed more routine system operation. A reduction in SO_2 removal efficiency was also noted over a period of time during which virgin MgO was the major part of the feed. Continued operations showed a recovery in efficiency when using regenerated MgO.

Performance tests for SO_2 removal run on recycled MgO gave an average 90% and 93.9% SO_2 removal efficiency with the scrubber operating at full rated capacity with SO_2 inlet concentrates of 771 PPM and 1418 PPM, respectively.

An explanation of the reduced efficiency when using virgin MgO is the more rapid formation of the complex, magnesium oxysulfate, with this more reactive material.

C) The co-current dryer system used in the PEPCO operation had a higher efficiency than the unit used in the oil-fired case. Extensive modifications of the counter-current dryer, to prevent material sticking to the walls,included elimination of most of the internal lifters. This would account for the higher co-current efficiency.

A number of instances were encountered during the PEPCO operations in which centrifuge cake feed was reported as sticking to the dryer walls. This was corrected several times "on the run" by reducing load. There were some cases where large deposits in the dryer resulted in a temporary shut-down.

It should be noted that the dryer had been designed to accept a set of chains, but these had not been installed pending the outcome of the development test work. Because of the unanticipated shut-down of the entire operation this potential was not demonstrated.

Dryer product from the PEPCO operation also proved more difficult to handle in this regeneration step, and resulted in a forced cutback in calciner feed rate to one half that obtained on the oil fired application. The reduced feed rate was correlated with a higher percentage of fines in the PEPCO dryer product, with only 22% as + 25 mesh in the PEPCO dryer product versus 58% in the same mesh size for the Boston Edison dryer product.

D) The wet fan performance and service was excellent. There were no signs of deterioration, wear or corrosion after all operations completion in either E.S.P. treated or untreated flue gas modes.

1.2 EFFECT OF PROCESS CHEMISTRY VARIATIONS ON SYSTEM PERFORMANCE

A) Production of magnesium sulfite hexahydrate rather than magnesium sulfite trihydrate resulted in little or no relief from the problems of "wet" centrifuge cake encountered in the oil fired application.

In normal operations, after attaining equilibrium, the centrifuge cake contained a mix of 82% as hexahydrate, 18% as trihydrate for its magnesium sulfite component. Materials handling problems were solved, as in the oil fired case, by designing the system to take any consistency of the centrifuge cake.

Improvements were made in redesigning the centrifuge discharge chute with a vertical wall on the solids discharge side of the machine and by lining the chute with plastic to overcome the sticking tendency of the cake.

B) Processing problems resulting from the formation of the complex SO_4Mg MgO H_2O have been described in A-2 above. Laboratory experiments indicate that this material is formed more readily when slaking a highly reactive magnesia (as virgin MgO grade used in this operation). Performance improves when the feed is composed of a high percentage of recycled MgO. It appears that the complex formation occurs during the step

$$MgO_{(s)} + H_2O\ (l) \longrightarrow Mg\ (OH)_2\ (s)$$

as subsequent laboratory tests showed no oxysulfate formation when the starting material was magnesium hydroxide.

C) Other Findings

Explanations were developed for the contradictions in the findings of various researchers on the thermographic analysis of the magnsium sulfite tri- and hexa-hydrates. It was shown that the dehydration can be characterized as one-step in an open, non-equilibrium process whereas in a closed, equilibrium process it proceeds at the expected two step dehydration. No evidence was found for the existance of any lower hydrate.

This work led to a simple and accurate method of analysis for the ratio of the two hydrates using thermogravemetric analysis (T.G.A.) techniques.

2.0 INTRODUCTION AND BACKGROUND

The Chemico-Basic magnesia slurry process for SO_2 control had been developed in laboratory and pilot plant studies over a number of years. The first prototype plant employing this system was intalled on an oil fired boiler at Boston Edison Co.'s Mystic Station on a program (New England SO_2 Control Project) that spanned four years from the signing of the contract to the completion of the test work in July 1974.

Concurrently with the development of the "New England Project" Chemico pursued other potential applications for this efficient flue gas desulfurization method.

Applications were investigated that would further demonstrate the broad range of its use, particularly the advantages in those areas where a regenerable, non sludge-producing sulfur oxides control method was mandated.

Simultaneously, forward thinking utilities, such as Potomac Electric Power Company, were also investigating a number of process schemes that would provide a means of control of their sulfur oxides emissions in order to satisfy the new limitations being imposed by municipal, state and Federal governments. A principal criterion in selection of the process for control of the recognized pollutants was the economics associated with the alternate of a low sulfur content fuel supply.

In late 1970 Chemico submitted a proposal to PEPCO for a prototype demonstration plant using the Chemico-Basic process. The prototype location was set for PEPCO's Dickerson generating station site approximately 35 miles northwest of Washington, D.C. in Montgomery County, Maryland.

This site consists of 1000 acres in a rural area (but with suburban expansion from Washington developing within a few miles of the plant) bordered on the west by the Potomac River and the Chesapeake and Ohio Canal National Historic Park. The site had also been planned by PEPCO as the location of a 1700 MW expansion of generating capacity which was to be coal fired in accordance with company considerations for a balance of fuel types. PEPCO had successfully developed methods for adequate disposal of its fly ash from the Dickerson Station; therefore, a major consideration for future planning was the selection of a process for sulfur oxides control which would be acceptable at this environmentally sensitive site. Regenerable processes were a major candidate because of the complexity that would be added to operation of the power plant by the requirement to dispose of the larger volume of residue that would be produced in the throw-away processes. The Chemico-Basic magnesia F.G.D. system was eventually selected by PEPCO and engineering for the plant to be built at Dickerson Station commenced in July 1971. Construction of the plant started in July 1972 and was completed in August 1973.

A limited test program had been planned for the PEPCO operation as it was hoped that the major operating problems would have been developed and corrected in operation at the first prototype in Boston by the time the Dickerson plant was in operation. However, a series of events including delayed start up of both the Boston Edison and the PEPCO project, unanticipated problems found on starting up the processes, and the lack of a regeneration facility for PEPCO's early runs resulted in a restriction on the operation of the plant. Those operations that were conducted evidenced areas requiring additional, intensive investigation and development.

Some of these were:

1) Identification of the cause and remedy of agglomerates in the MgO slurry system.
2) Determination of losses and loss points as discrepancies in production and consumption calculations pointed to leaks of large magnitude in the magnesia system.
3) Determination of the extent and rate of corrosion as samples obtained from rail cars of $MgSO_3$ shipped to Rumford showed much higher levels of iron contamination than any experienced in the Boston Edison Operation.
4) Establishing actual SO_2 and particle removal efficiencies.

In order to provide information and solutions to these problems, a more comprehensive development and test program was prepared by Chemico and accepted by PEPCO in June 1974.

A major factor in planning of the new program was the extension of the data gathering and interpretation methods that had been used for the successful operations at Mystic station to operations of Dickerson. This immediately made the large data and correlation file which had been developed during the previous program accessible for extension and comparison with data generated from the coal fired application.

The program was divided into a series of tasks and phases in order to assure that the development proceeded in a fashion that would establish priorities first, attaining the fundamental requirements, and second leading to demonstration of the system reliability, if time permitted.

The first phase of the program was to determine control parameters that would allow sustained plant operations under conditions that established system equilibrium. Complications were encountered during this period because of need to continue shake-down operations with new and modified equipment.

The tasks slated for completion during this first phase were:

1) Evaluation of changes to the MgO slurry system.
2) Determination of the complete material balance at the plant.
3) Measurement of system pH throughout the plant.
4) Preliminary measurements of SO_2 removal.
5) Preliminary operation with regenerated MgO from the calcining facility.
6) Evaluation of equilibrium conditions in the fly ash system with the precipitator in operation.
7) Coordination of operations during the preliminary system testing (conducted by an outside testing service) in order to provide initial SO_2, SO_3 and particle control data.

The second phase in the program had as its goal the optimization of operation of the system at the Dickerson site and determination of the process conditions at the regeneration facility necessary to provide an alkali suitable for reuse after several cycles. Some differences between the oil and coal fired applications could be expected resulting from coal ash and/or first stage particle removal. These differences would affect the:

1) Efficiency of SO_2 removal.

2) Degree and rate of oxidation of the salts formed.

3) Efficiency of regeneration of MgO.

4) Activity of the regenerated MgO.

Evaluation of these areas required an investigation of the degree of contamination of MgO slurry with unremoved particulate matter and assessment of the SO_2 removal efficiency of each stage for various cases. This was to be done concurrently with the determination of the effect of fly ash contamination on the regenerated MgO and acid plant in long term operations.

In order to achieve these results in the shortest term possible, extensive use was made of the data file complied from the operating and analytical information logged during the Boston Edison operations. Using data processing methods, conditions at the Dickerson plant could be quickly analyzed and compared with information in the file. In this way trends indicating potential operating or process difficulties could be rapidly identified and the necessary corrective action taken so that maximum utilization of the intergrated plants would be achieved. The plan involved investigation of several parameters, such as system pH, pressure drop, recycle and gas flow and the resultant ash concentrations and solids levels in the various streams, dryer temperature, and liquid flow rates.

Most variable excursions would occur as a natural effect of process deviations. With data recording, sampling and chemical analysis on a routine schedule, they would be measured and their effects evaluated. Some of the variables such as gas flowand recycle rates were to be artifically varied in order to reduce the danger of gross contamination of the absorbent before the influence of the build-up of inerts on the process could be assessed.

In order to determine the effect of changes in these parameters a factorial experimental program was to be used where possible.

It can be anticipated that the widest excursions in operating conditions would occur during the earliest part of the program as the operators gain skill in running the plant and as deliberate changes in conditions are made to overcome process and mechanical problems. As operations continue, these variations diminish or disappear as greater control is gained over the system so that steady conditions will be experienced during the latter part of this period.

Phase three of the program was to be a time extension so that continuous operations under the optimized conditions developed during Part II could be achieved. The duration of this phase would be contingent upon the time required to achieve three additional cycles of the inventory. Overlap with the previous period would result in a total of five regenerations of the magnesia and replacement of 1/2 of the inventory. This phase was never implemented.

Program Schedule

Phase I of the program, the conclusion of the shakedown operations and the preliminary operational testing, was estimated to require a one to two month period for completion. System feasibility as well as the suitability of the installed equipment was to be demonstrated in this Phase I. With this information plant design and equipment selection could be undertaken for any similar installation.

The second phase of the program, investigation of varying operating conditions on performance for system optimization and the determination of the quality of regenerated magnesia to obtain the required levels of SO_2 control was estimated to require an additional four months. At the conclusion of the testing program the detailed information on operating values were to be used to fix control parameters for the final phase.

Phase three, a continuous operating period of three to four months' duration was to conclude the program.

A measure of the success of the program that was achieved stems from the reorientation of the program from the limited process guarantee tests to the more expansive one described in the preceding sections. In order to accomplish those goals, however, it was necessary to expand the socpe of work that was being undertaken at the Dickerson site. Provision had to be made for a staff of chemists and technicians to provide the analytical and testing support, technicians and operating engineers were necessary to provide supervision and advisory services for the plant operations and an overall organization had to be created to direct the project, assimilate the data, implement the changes necessary to solve the system problems and to coordinate the operations being conducted at both Dickerson and Rumford.

This was accomplished through a joint agreement between PEPCO, Chemico and the Environmental Protection Agency in which PEPCO provided the funding for the bulk of the operations at both plants, Chemico provided the project management function and EPA funded the analytical program and its support. EPA also provided funding for other emissions testing programs conducted to determine the effect of transient conditions.

3.0 PROCESS DESCRIPTION

The pollution abatement system installed at PEPCO's coal burning Dickerson Station on the Number 3 boiler is composed of a particle control scrubber and an SO_2 control absorber, both contained in a single, two stage unit (Fig 1). Particle control is effected by use of a Chemico designed variable throat venturi as a first stage followed by a venturi absorber using the Chemico-Basic Magnesium Oxide System for Flue Gas Desulfurization (FGD) as the second stage. The particle control stage, including the recycle and slurry streams and fly ash thickening units are independent and separate from the SO_2 control stage, thus preventing fly ash contamination of the magnesia used as the absorbent in that stage. Both the particle control streams and the F.G.D. streams are operated in a closed loop mode.

The original system concept was to provide a prototype demonstration unit designed to process one-half of the flue gas from the Unit 3 boiler, as shown in Fig. 2. In one configuration the flue gas to be treated is diverted from the stack after the electrostatic precipitator by using the venturi scrubber induced draft fan. A second configuration allows the ESP to be bypassed through new ductwork so that tests could be conducted with the full particle loading from the coal fired boiler treated in the wet scrubber unit. In either case, the flue gas cleaned in the first stage was processed for SO_2 removal in the second stage.

In the Chemico-Basic process for sulfur dioxide removal from flue gas the SO_2 is absorbed in a slurry containing

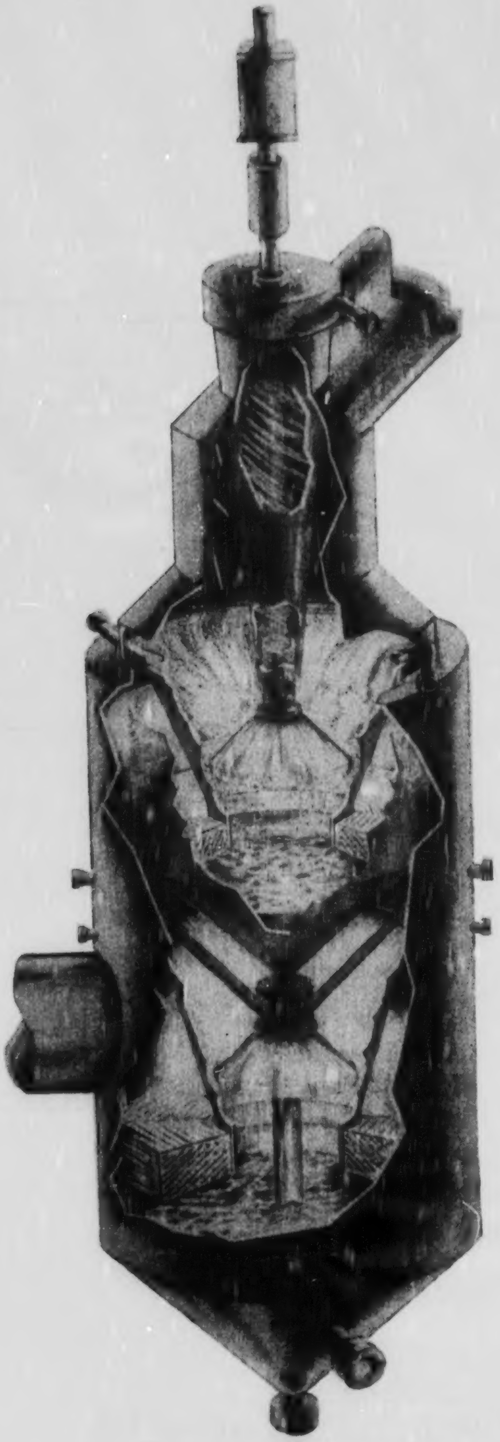


FIGURE 1

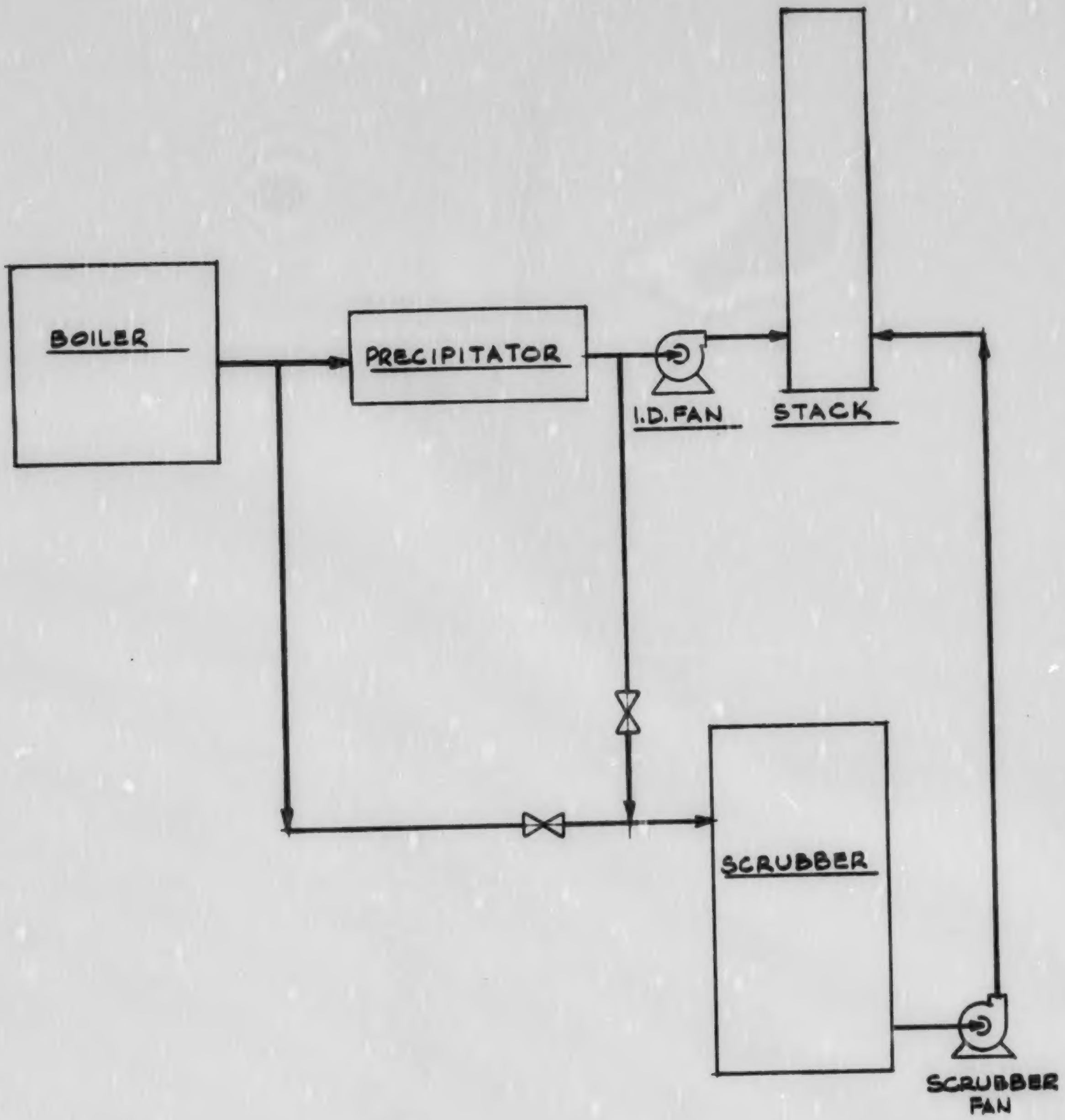

BOILER
PRECIPITATOR
I.D. FAN
STACK
SCRUBBER
SCRUBBER FAN

FIGURE 2
FGD GAS FLOW SCHEMATIC
DICKERSON No 3 - PEPCO

magnesia, precipitating the solid magnesium sulfite. The magnesium sulfite, after appropriate steps to reduce its moisture content, is then processed at a separate installation to regenerate the magnesia and recover the SO_2 for use in chemical manufacture. In this case the regeneration was accomplished at the E.P.A. facility at Essex Chemicals' Rumford Acid Plant in Rhode Island which had been built for the New England SO_2 Control Project, with the recovered SO_2 used in the manufacture of sulfuric acid.

3.0.1 Facility Description

Dickerson Station is located on the Potomac River outside the town of Dickerson, Maryland. The plant is situated in a rural, non industrialized area about 35 miles northwest of Washington, D.C. Coal is delivered to the plant by rail.

The station has three electric generators each rated at 190 MW. A fourth generator, rated at 800 MW is scheduled for installation nearby on the plant site by 1982. The installed 95 MW FGD system is sized to handle approximately one-half the exhaust gas flow from Unit No. 3. Pertinent data on this Unit are given in Table 1. Unit No. 3 has a dry-bottom coal-fired boiler that was designed by Combustion Engineering and installed in 1962.

The coal presently burned has an average gross heating value of 11,700 BTU/lb. Average ash and sulfur contents are 14 percent and 2 percent, respectively.

The boiler is fitted with an electrostatic precipitator (ESP) designed and installed by Research-Cottrell in 1962. Particle collection efficiency was estimated to be 94 percent. The FGD system is installed so that it can receive exhaust gas either from the outlet or breeching ahead of the ESP.

TABLE 1

PERTINENT DATA ON PLANT DESIGN, OPERATION AND ATMOSPHERIC EMISSIONS

Boiler data - Dickerson No. 3 - PEPCO

Rated Generating Capacity MW	190
Average Capacity Factor (1974), %	81
Boiler Manufacturer	Combustion Engineering
Year Placed In Service	1962
Unit Heat Rate, BTU/kwh	9180
Maximum Coal Consumption, Ton/Hr.	74.5
Maximum Heat Input, BTU/hr.	1744×10^6
Stack Height Above Grade, Ft.	400
Flue Gas: Design Rate, ACFM	450,000
Maximum Rate, ACFM	650,000
Flue Gas Temperature, °F	245
Emission Controls:	
Particulate Matter	ESP with a venturi scrubber on half of gas flow.
SO_2	Venturi absorber on half of the gas flow.
Particle Emission Rate:	
Allowable, g/scf dry	0.03
SO_2 Emission Rates:	
Allowable, lb/Million BTU	1.6(1)

---

(1) 1% sulfur coal equivalent

3.0.2 Pollution Abatement System Description

A schematic representation of the pollution abatement process installed at the power plant is shown in Fig. 3. The principal element of the system is the two stage venturi scrubber/absorber which combines in a single vessel both particle removal and SO_2 control. Each of the two functions are separated in the unit with the flue gas being treated sequentially, first to remove fly ash then to reduce its SO_2 content to acceptable levels. Besides providing the means of cleaning the flue gas, this sequence assures that a negligible amount of fly ash will be captured in the magnesia slurry used for SO_2 removal in the second stage. This is necessary in order to prevent a buildup of coal fly ash which would result in the contamination of the regenerated MgO. A further advantage of this design is the ability to use the particle removal stage independently should it be desirable to continue the unit in operation without SO_2 control.

The Chemico/Basic's Magnesium Oxide System for the recovery of sulfur dioxide from plant flue gases utilizes the sulfur dioxide absorption characteristics of an aqueous slurry of magnesium sulfite, magnesium sulfate, and magnesium oxide. The process is accomplished in several steps:

Absorption
Dewatering & Drying
Materials Handling
Calcination

with the regeneration/recovery step (calcination) performed at a remote site.

Absorption

In the absorption step the process chemistry which describes the removal of SO_2 from the flue gas can be most

THE DICKERSON REGENERATIVE WET SCRUBBING SYSTEM

POTOMAC ELECTRIC POWER CO. PROTOTYPE

PRECIPITATOR/SCRUBBER - ABSORBER

MgO ADDITIVE SYSTEM FOR SO_2 RECOVERY

SCHEMATIC PROCESS FLOW SHEET

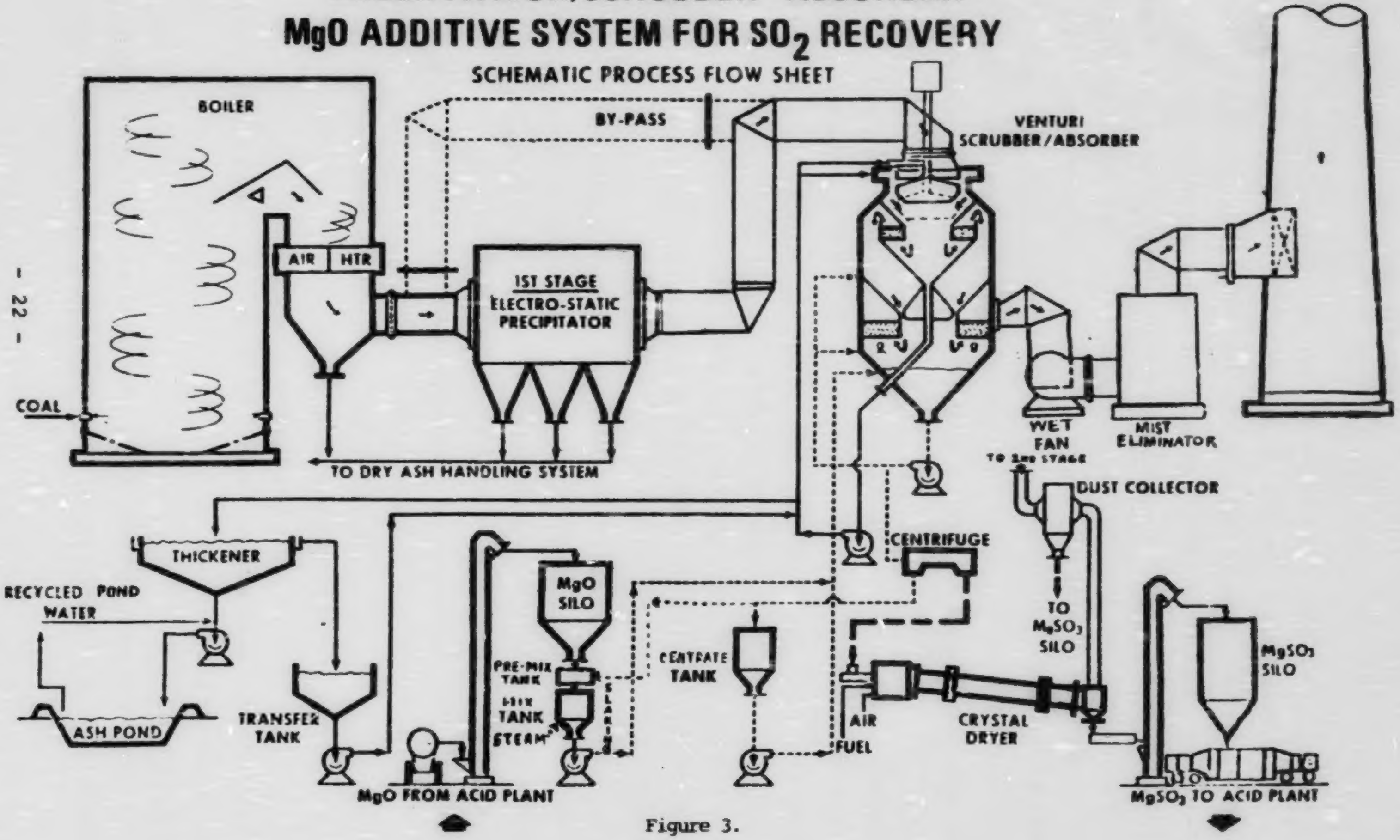

BOILER
BY-PASS
VENTURI SCRUBBER/ABSORBER
AIR
HTR
1ST STAGE ELECTRO-STATIC PRECIPITATOR
COAL
TO DRY ASH HANDLING SYSTEM
WET FAN
MIST ELIMINATOR
TO 2ND STAGE
DUST COLLECTOR
THICKENER
RECYCLED POND WATER
ASH POND
TRANSFER TANK
CENTRIFUGE
MgO SILO
PRE-MIX TANK
MIX TANK
STEAM
SLURRY
CENTRATE TANK
TO $MgSO_3$ SILO
$MgSO_3$ SILO
AIR
FUEL
CRYSTAL DRYER
MgO FROM ACID PLANT
$MgSO_3$ TO ACID PLANT

Figure 3.

simply explained as the diffusion of SO_2 through the flue gas to a liquid surface, then absorption of the SO_2 with the hydrated form of MgO, i.e.

$$MgO_{(aq)} + SO_2 \longrightarrow MgSO_3 \quad (3-1)$$

The $MgSO_3$ produced has a low solubility and can be separated from the absorbing slurry as a solid.

In the process, the flue gas containing sulfur oxides enters a venturi absorber of special design, and contacts the absorbing media which is an aqueous slurry of magnesium oxide, magnesium sulfite, magnesium sulfate, and a small percentage of other components. The process of SO_2 removal that occurs is explained by conventional mass transfer principles.

The venturi absorber can be considered as similar to a co-current, packed vessel. In the venturi the liquid slurry is introduced and flows downward on a surface over which passes an accelerating gas stream. The high velocity gas stream flowing over the liquid causes wave motion on the fluid surface, the waves increase in amplitude and finally disperse as fine droplets. Thus, the whole mass of liquid can be distributed in the form of atomized droplets in the gas stream.

There are several advantages in using the venturi as an absorber. The absorption surface is dispersed into and flows with the gas stream during the time absorption is occuring, thus eliminating the problems of plugging associated with conventional packed towers (the surface area per unit volume is approximately equivalent to dumped 3 inch rachig rings). Due to system dynamics, this surface area relation is relatively invariant over wide turn-down ratios, and high removal efficiencies can be maintained over the normal operating range of the power plant's boiler.

Dewatering & Drying

A stream from the absorption system enters a centrifuge where the solids which were formed by the absorption reaction in the slurry are separated. This bleed stream is controlled in order to maintain a constant solids content in the recycle slurry by removal of product magnesium sulfite and unreacted magnesium oxide and any precipitated magnesium sulfate. The system is operated so that the absorbed SO_2 is removed as an equivalent amount of the magnesium sulfur compounds.

The wet centrifuge cake containing hydrated magnesium sulfite, magnesium oxide and magnesium sulfate plus other solids removed in the venturi absorber-centrifuge system is passed to a rotary dryer to remove both unbound water and most of the water of crystallization. The dry product is easy to store, and the removal of water reduces shipping costs.

Materials Handling

The anhydrous magnesium sulfite and magnesium sulfate produced in the dryer is conveyed to a storage silo for transportation by rail or truck to the recovery acid plant. The same transport is used to return regnerated magnesia to the magnesium oxide silo at the power plant on the return trip.

Calcination System

Calcination is the process used for regeneration. The magnesium sulfite which has been separated and dried is thermally decomposed as represented by the following reaction:

$$MgSO_3 \longrightarrow MgO + SO_2 \qquad (3\text{-}2)$$

The recovered SO_2 is used in the production of sulfuric acid and the regenerated MgO returned to the process for reuse. The dry product transported to the regeneration acid plant from the power plant is received, weighed and pneumatically conveyed to a storage silo. It is fed to a direct-fired rotary kiln at a metered rate, and calcined to both generate sulfur dioxide gas and regenerate magnesium oxide. Coke can be added to provide a reducing atmosphere, as necessary, to reduce the residual magnesium sulfate to the oxide and sulfur dioxide. The hot flue gas containing sulfur dioxide and dust enters a cyclone where essentially all the dust is removed and returned to the calciner. The flue gas then enters a venturi scrubber for final dust cleaning. At the same time, the gas is cooled and adiabatically saturated.

The saturated flue gas is further cooled in a direct contact packed tower to meet the requirements of the acid plant water balance. The cleaned flue gas then enters the drying tower of an existing 50 T/D acid plant. At the Essex Chemicals' installation, the resultant product made from the recovery of the sulfur dioxide is 98% sulfuric acid.

The regenerated magnesia is cooled, conveyed to the magnesia storage silo, and recycled back to the power plant.

3.1 DETAILED PROCESS DESCRIPTION

The process flow sheet for the facility installed at Dickerson is shown in Fig. 4, material balances are given in Table 2 with the description of each item given in the Equipment List, Table 3.

3.1.1 Particle Removal

The first stage of the system uses a Chemico variable throat venturi. In this design, a 60° tapered plug is suspended by a height adjustment mechanism in a 60° converging section of the scrubber. By stroking the mechanism through its entire adjustment the throat opening can be varied allowing up to a two foot maximum spacing. In normal operation, with standard liquor circulation rates, the throat width has been adjusted to maintain an eleven inch pressure difference across the particle removal stage.

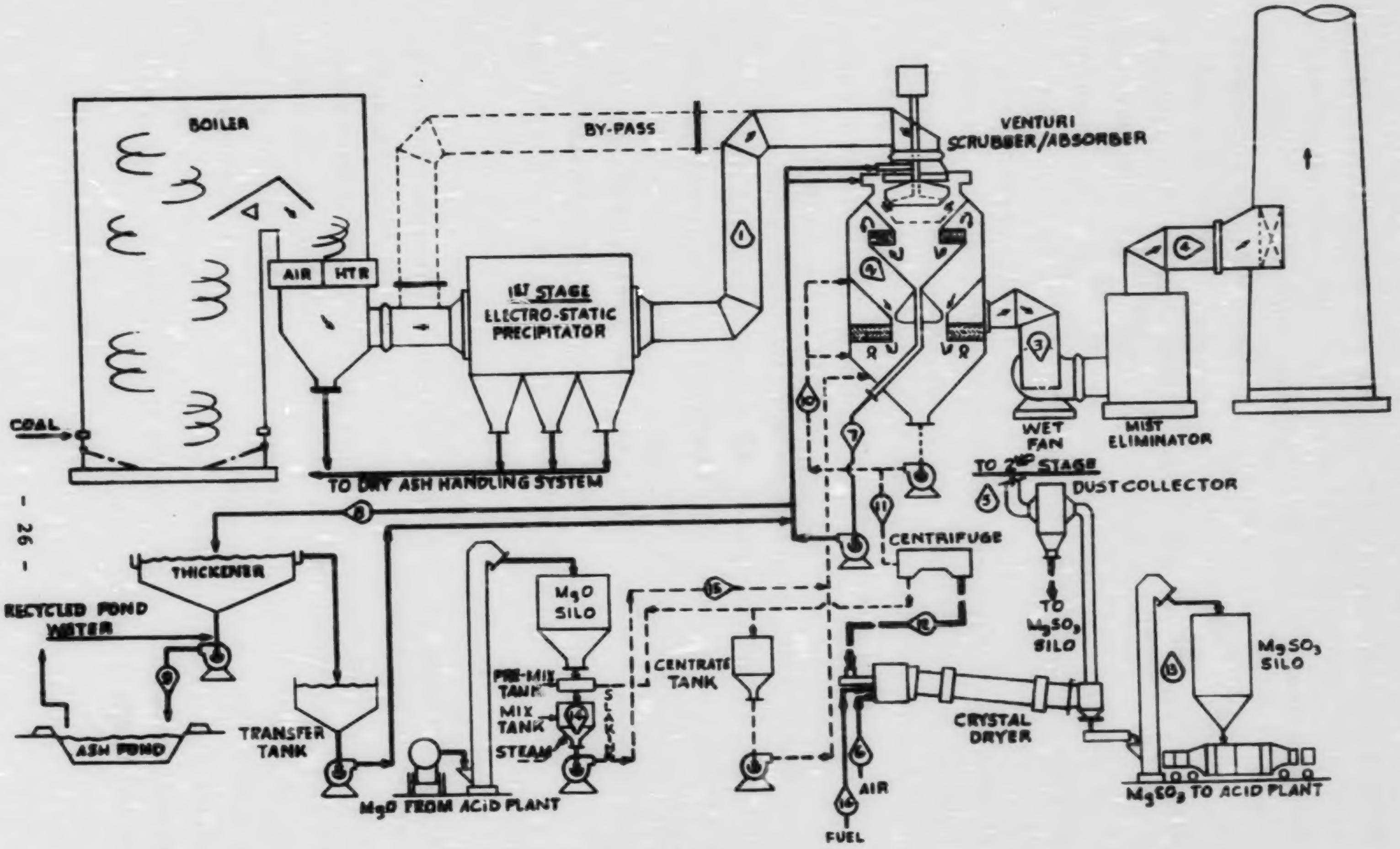

BOILER
COAL
AIR
HTR
BY-PASS
1ST STAGE
ELECTRO-STATIC
PRECIPITATOR
TO DRY ASH HANDLING SYSTEM
VENTURI
SCRUBBER/ABSORBER
WET
FAN
MIST
ELIMINATOR
TO 2ND STAGE
DUST COLLECTOR
TO
MgSO3
SILO
CENTRIFUGE
THICKENER
RECYCLED POND
WATER
ASH POND
TRANSFER
TANK
MgO FROM ACID PLANT
MgO
SILO
PRE-MIX
TANK
MIX
TANK
STEAM
SLAKER
CENTRATE
TANK
CRYSTAL
DRYER
AIR
FUEL
MgSO3
SILO
MgSO3 TO ACID PLANT
1
2
3
4
5
6
7
8
9
10
11
12
13
14
15
16

FIGURE 4
PEPCO REGENERATIVE
WET SCRUBBING SYSTEM

TABLE 1

STREAM PROPERTIES AND COMPOSITION

SO_2 AND FLY ASH REMOVAL SYSTEM

STREAM NUMBERS		7	8	9	10	11	12	13	14	15	16
		LIQUID AND SOLID STREAMS									
Temp.	°F	117	117							38.1	
Flow	GPM	5,165	980	140	10,621	170				13.7	
Solids Content,%		2.0	2.0	13.2	13.3	13.2					
H_2O	Lb/Min	42,870	8,166	1,102	82,850	1,322	10.4			300	
$MgSO_3$	Lb/Min							86.3			
$MgSO_4$	Lb/Min							4.9			
$MgSO_3 \cdot 6H_2O$	Lb/Min					178	178	2.0			
$MgSO_4 \cdot 7H_2O$	Lb/Min					9.9	9.9				
MgO	Lb/Min					9.9	9.9	9.9	45.3	45.3	
Inerts	Lb/Min					2.4	2.4	2.4	2.4	2.4	
Fly Ash	Lb/Min	875	166	166		0	0	0			
Total	Lb/Min	43,745	8,332	1,268	95,520	1,522	210.6	105.5	47.7	347.7	
Fuel	Btu/Hr										16.6×10^6

STREAM NUMBERS		1	2	3	4	5	6
		GAS STREAMS					
Temp.	°F	259	117	121	121	450	70
Pressure	In. W. G.	-11	-26	-36	0	-5.5	-0.6
Flow	ACFM	295,000	263,000	286,000	260,000	21,600	10,200
SO_2	Lb/Min	62.5	62.5	6.25	6.25		
H_2O	Lb/Min	581	1,209	1,420	1,420		
Dry Gas	Lb/Min	15,820	15,820	16,457	16,457		
Particulate	Lb/Min	166.6	0.85	0.89	0.89	0.22	

TABLE 3

EQUIPMENT LIST

Item Description	Code No.	Quantity	Motor Horsepower (HP)
MgO Make-Up Tank	G-101	1	
Dust Collector-MgO Silo	G-102	1	
Thickener	G-201 A,B	2	
Transfer Tank	G-202	1	
Distribution Box	G-203	1	
Sump Tank	G-204	1	
Mother Liquor Tank	G-301	1	
Dust Collector-Dryer	G-401	1	
MgO Storage Silo	I-101	1	
$MgSO_3$ Silo	I-401	1	
MgO Feed Pump	J-101 A,B	2	10
First Stage Recycle Pump	J-201 A,B	2	350
Second Stage Recycle Pump	J-202 A,B,C	3	250
Underflow Pump	J-203 A,B,C,D	4	
Transfer Pump	J-204 A,B	2	125
Sump Pump	J-205	1	30
Mother Liquor Pump	J-301 A,B	2	20
Induced Draft Fan	K-201	1	3500
MgO Agitator	M-101	1	5
Mother Liquor Agitator	M-301	1	3
MgO Weigh Feeder	O-101	1	
Dryer Feed Conveyor	O-401	1	1
Dryer Discharge Conveyor	O-402	1	2
Dryer Discharge Elevator	O-403	1	2
$MgSO_3$ Conveyor	O-404	1	2
$MgSO_3$ Weigh Feeder	O-405	1	
Two-Stage Venturi Scrubber	R-201	1	
Mist Eliminator	R-202	1	
Variable Speed Coupling	R-203	1	

TABLE 3 (Cont'd)

EQUIPMENT LIST

Item Description	Code No.	Quantity	Motor Horsepower (HP)
Centrifuge	R-301	1	200
Rotary Dryer	R-401	1	40
Dryer I.D. Fan	K-402	1	100
Instrument Air Dryer	V-101	1	

Liquor is circulated in this particle removal stage by recycle pumps, J-201 A & B (one operating, one spare, each with a design capacity of 5700 GPM and driven by a 350 HP motor) through rubber lined distribution piping to the upper level of the scrubber R-201. The recycle flow is split into two streams before reentering the scrubber. The flow in each stream is adjusted and set by pinch valves to provide uniform irrigation of the converging surfaces. The outer converging section is supplied by ten 4" tangential nozzles while the adjustable cone is supplied by three 6" lines feeding a single bull nozzle.

Flue gas enters the vessel and passes downwards through the irrigated converging section. As the gas is accelerated through the throat the liquid film is atomized and distributed through the gas stream. Particle collection occurs on the dispersed droplets through the mechanisms of impaction, diffusion and attraction. The flue gas and dispersed liquor continue through the central down-comer to the basin area of the first stage where the gas turns 180° upwards to pass through the first stage mist eliminators. The liquid droplets are first disengaged in the turning zone and fall into the first stage liquor pool, droplets small enough to be carried with the gas stream are captured by the slot type mist eliminator. The cleaned gas continues out of the first stage through the annular space between the outer shell and the wall of the first stage reservoir.

A bleed stream, about 15% of the total flow from the recycle pumps, is sent to a distribution box and thence to one of two thickeners, G-201 A & B, each of these is 40' in diameter with a rake mechanism feeding a central underflow discharge. The thickened underflow, approximately 20 GPM at 40 percent solids, is discharged to a distribution tank where water from the lower pond is added and the mixture is pumped to the upper settling pond. The pond overflow cascades through a total of four settling ponds in series and the clear water from the lowest is again returned to the dilution tank.

The thickener overflow drops into a transfer tank and is returned to the first stage recycle loop by transfer pumps J-204 A & B, (one operating, one spare, each designed for 1190 GPM and driven by a 125 HP motor) so that closed loop operation is maintained.

It should be noted that the mechanisms for collection of particulate matter or sulfur dioxide both depend on the formation of droplets dispersed in the flowing gas. In the Chemico venturi the mechanism of atomizing the liquid by the accelerated gas stream is the same for either stage. However, the mode of operation of the stages is different. In the case of particle collection a specific droplet size distribution will be found to provide the maximum collection efficiency and the size distribution is directly related to a gas velocity flowing over the liquid surfaces. In the case of gas adsorption it is necessary to maintain a uniform specific surface area (i.e. area of droplet surface per volume of gas) which in the venturi, is self regulating, after initial adjustment, over a range of gas flows. Thus, first stage efficiency is maintained by operation at a fixed pressure drop to keep the best droplet size distribution while second stage efficiency is controlled by operation at fixed liquor recycle rates to maintain a uniform surface area of the dispersed droplets.

3.1.2 Detailed Process Description - SO_2 Removal

In this system magnesia, both regenerated and make-up material, is transferred from pneumatic discharge hopper trucks to the elevated MgO storage bin through a 4 in. pneumatic unloading system. The MgO bin, I-101, is 25 ft.

in diameter and 46.5 ft high. Magnesia is fed from the silo, which is equipped with a vibrating hopper bottom, O-103, to a 3760 gallon capacity MgO make-up tank G-101 by an adjustable weigh feeder, O-101, with mother liquor to make the desired MgO slurry composition. A small pre-mix tank is interposed between the weighing system and the steam heated make-up tank to insure dispersion of the MgO powder and to act as a vapor seal. Heated magnesia slurry is added to the recycle stream by the MgO make-up pump, J-101 A, B. The magnesia slurry addition rate is controlled by the operator to maintain the pH of the absorbing slurry at the desired value.

The recycle stream for the absorber is circulated at a rate of 10,767 GPM by pumps, J-202 A, B and C, (the capacity of each of these 250 HP pumps is half the design flow of the recycle stream) to provide a slurry dispersion within the vessel sufficient for the desired SO_2 removal.

The recycle stream itself is split into two streams for distribution before reentering the venturi absorber. The flow in each stream is adjusted and set by plug valves to provide a uniform irrigation to the converging surfaces. Distribution of absorbing slurry is made by ten, six inch tangential nozzles supplying the outer converging section while the inner cone is supplied by a single central nozzle.

The slurry enters the upper part of the absorber stage with the flue gas which is pulled through the system by the I.D. fan K-201 driven by an 3500 HP motor, designed to handle 286,000 ACFM at a discharge head of -26 in. water. The gas and slurry mixture passes through the throat area into a diverging section, then into a central downcomer. To exit the vessel the flow of the cleaned flue gas turns one hundred eighty degrees upwards. In this step most of the larger liquid droplets are disengaged from the flowing gas stream and fall to the slurry pool in the conical base of the absorber. The treated flue gas continues upwards through a slot-type mist eliminator to further remove any entrained liquid before the gas exits to wet fan K-201. The fan wheel of K-201, driven through a variable speed drive, is continuously sprayed with water to prevent solids build up on the blades. Entrained liquid in the gas stream leaving the fan is removed in Mist Eliminator vessel R-202 before passing the stack.

A stream of slurry, approximately 1.6% of the total recycle, is taken from the discharge of the recycle pumps to the centrifuge, R-301. This is a 36 in. x 72 in. horizontal, solid bowl unit driven by a 200 HP motor, and in normal operations it removes 50% of the solids in the liquid stream going to it.

The solids separated in the centrifuge contain an amount of SO_2 (as $MgSO_3$ or $MgSO_4$) equivalent to the amount removed in the absorber. The centrate is discharged directly to an agitated tank, G-301 of 1375 gallon capacity which serves as a pump tank for this system, and then is returned to the MgO slurrying system by pumps J-301 A & B. The basin of the absorber is the reservoir for the system and contains approximately 20,000 gallons of slurry. The other sources of water entering the system are small amounts added as pump seal water and an additional quantity used as a mist eliminator wash; the latter is an intermittent addition.

The centrifuge cake separated in the centrifuge is fed to a dryer by screw conveyor O-401, a 35 ft. long unit with a 12 in. diameter ribbon flight. The dryer is a rotary unit 8 ft. in diameter by 50 ft. long. Heat is supplied from an integral 18 ft. long oil fired combustion chamber at a maximum heat release rate of 26.3×10^6 BTU/hr.

The dryer off-gas exits in to a cyclone, G-401, designed for 97% removal of the solids entering with the gas. From the cyclone dust collector, the gas discharges via a booster fan, K-402, to the venturi absorber where, mixed with the entering flue gas, it is cleaned of its remaining particle load. The separated fines collected by the cyclone returns directly to the dry product pneumatic conveying system discharging to the product silo.

The dryer product is discharged to the dryer discharge conveyor, D-402, 21.75 feet long unit with a 9 inch diameter solid flight screw. It then exits into a $MgSO_3$ weigh feeder O-405 at the boot of the $MgSO_3$ discharge elevator, O-403, which is designed to handle 5 tons/hour of dried magnesium sulfite. The product then travels through another conveyor, O-404, located on top of the $MgSO_3$ storage silo. The storage silo is 66.5 feet high and 25 feet in diameter, and is equipped with a vibrating discharge hopper, O-406.

The product $MgSO_3$ is finally gravity loaded into a waiting hopper truck for shipment to the MgO regeneration plant.

3.2 DETAILED PROCESS DESCRIPTION - MAGNESIA REGENERATION SYSTEM

Magnesium sulfite from the absorption system is received at the regeneration facility. The regeneration section flow sheet is shown in Fig 5 . Material balances are given in Table 4. The salt is unloaded from hopper trucks to the sulfite silo I-401 through a 4 in. pneumatic conveying system similar to that used for MgO at the power station. The sulfite silo was an existing tile structure, 25 ft. in diameter x 45 ft. high, which was water-proofed before this use. Sulfite is unlaoded from this silo through a pair of inclined screw conveyors, $MgSO_3$ conveyors 1 and 2, O-509 and O-510, equipped with 12 in. diameter helicoid flights, which carry the material from the below grade discharge spout to the boot of the $MgSO_3$ elevator.

This elevator, O-511, a continuous discharge unit 54 ft. high, is designed to handle 9 TPH of $MgSO_3$. It discharges

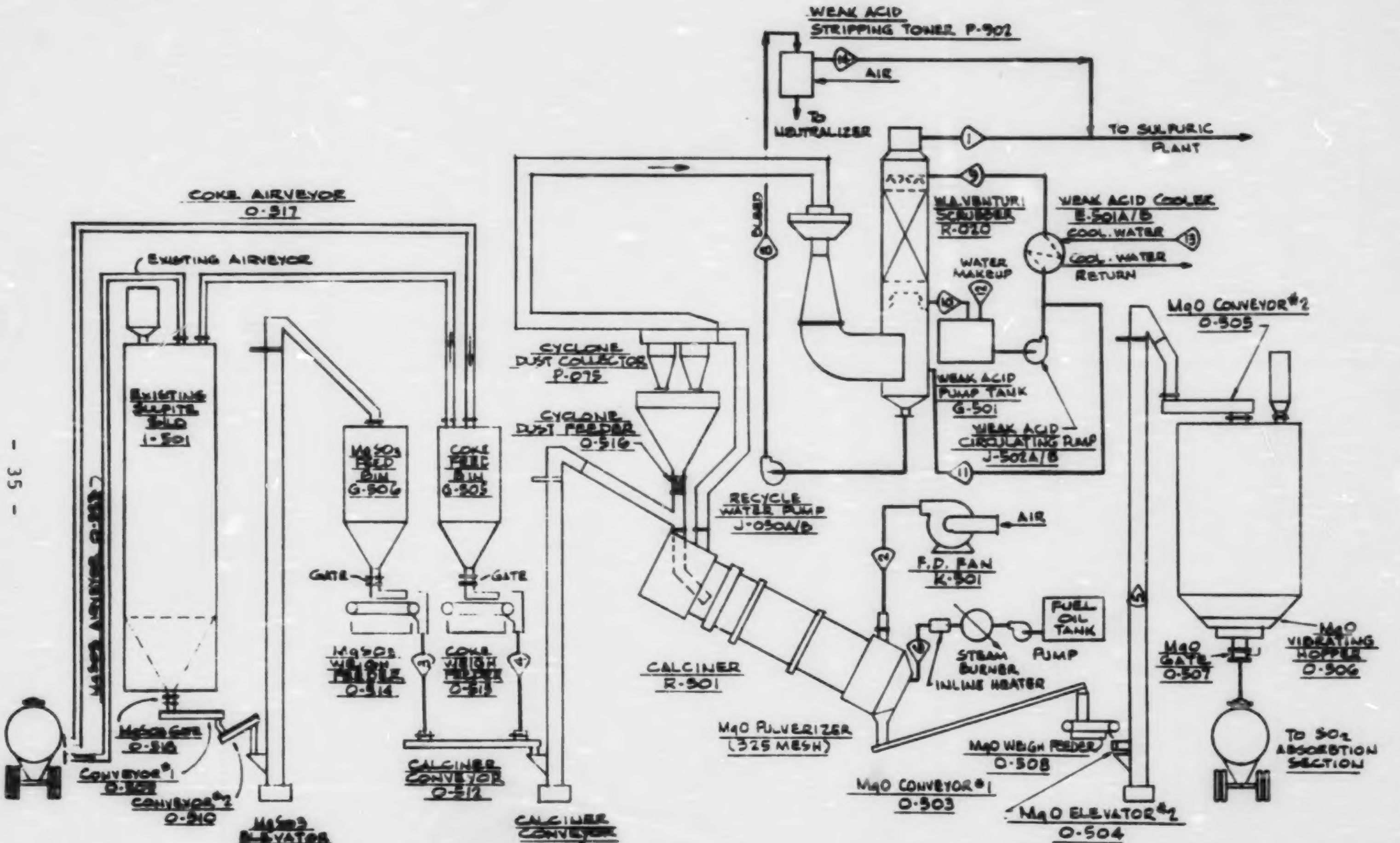

COKE AIRVEYOR O-517
EXISTING AIRVEYOR
EXISTING SULFITE SILO I-501
MgSO3 AIRVEYOR
MgSO3 GATE O-518
CONVEYOR #1 O-509
CONVEYOR #2 O-510
MgSO3 ELEVATOR O-511
MgSO3 FEED BIN G-506
COKE FEED BIN G-505
GATE
GATE
MgSO3 WEIGH FEEDER O-514
COKE WEIGH FEEDER O-513
CALCINER CONVEYOR O-512
CALCINER CONVEYOR O-515
CYCLONE DUST COLLECTOR P-075
CYCLONE DUST FEEDER O-516
CALCINER R-501
RECYCLE WATER PUMP J-050A/B
BLEED
WEAK ACID STRIPPING TOWER P-902
AIR
TO NEUTRALIZER
TO SULFURIC PLANT
W.A. VENTURI SCRUBBER R-020
WEAK ACID COOLER E-501A/B
COOL. WATER
COOL. WATER RETURN
WATER MAKEUP
WEAK ACID PUMP TANK G-501
WEAK ACID CIRCULATING PUMP J-502A/B
AIR
F.D. FAN K-501
FUEL OIL TANK
PUMP
STEAM BURNER
INLINE HEATER
MgO PULVERIZER (325 MESH)
MgO WEIGH FEEDER O-508
MgO CONVEYOR #1 O-503
MgO ELEVATOR #2 O-504
MgO CONVEYOR #2 O-505
MgO GATE O-507
MgO VIBRATING HOPPER O-906
TO SO2 ABSORBTION SECTION
1
5
10
11
13
14
17
18

FIGURE 5

MgO REGENERATION SYSTEM

TABLE 4

STREAM PROPERTIES AND COMPOSITION
MgO REGENERATION SYSTEM

STREAM NUMBERS		3	4	5	6	7	8	9	10	11	12	13
		LIQUID & SOLID STREAMS										
Temp	°F			300	250	160	160	90	150	150	Amb.	80
Pressure	PSIG											
Flow	GPM				1.5	91	2	119	125.5	80	1.5	205
$Mg(HSO_3)_2$	lb/Min						0.88					
MgO	lb/Min	8.24		38.35								
$MgSO_3$	lb/Min	72.50										
$MgSO_4$	lb/Min	4.12					1.5					
Inerts	lb/Min	1.89		1.81			0.1					
Fly Ash	lb/Min	0.25		0.25								
Total Solids	lb/Min	87.0	2	40.41			0.1					
H_2O	lb/Min						13.4					
Total Flow	lb/Min	87.0	2	40.41		760	15.88	99.2	1045	65	12	1708
Fuel Oil	lb/Min				12							

STREAM NUMBERS		1	2	14
		GAS STREAMS		
Temp.	°F	100	70	100
Pressure	In. W. C.	-45	0	-45
Flow	ACFM	2,980	2,140	640
Flow	SCFM	2,460	2,100	529
Total Dry Gas	lb/Min	206.2	161	39.0
H_2O Vapor	lb/Min	9.0		0.8
Total Wet Gas	lb/Min	215.2		39.8
SO_2	lb/Min	44.7		
O_2	lb/Min	1.8		
Dry Gas MW		34.4	29	29

to the $MgSO_3$ feed bin, G-506, a 6 ft. diameter by 11 ft. high vessel equipped with a vibrating bottom, which also serves as a surge bin for the weigh feeder O-514. This weigh feeder is continuously variable to a maximum capacity of 3.6 TPH and discharges to the calciner conveyor O-512, a 33 ft. long horizontal unit equipped with a 9 in. diameter helicoid screw. Coke is also discharged to this screw which serves to mix the two components as calciner feed prior to processing.

The coke is pneumatically conveyed to the coke feed bin G-505 9 ft. diameter x 19 ft. high, which also serves as the storage bin for this material. It is metered to the calciner conveyor by a weigh feeder which is also continuously variable up to a maximum capacity of 2.5 pounds per minute. The two streams (3 and 4 of Fig. 5) enter at points 9 feet apart and mix in the remaining 22 ft. section of the conveyor before reaching the calciner elevator O-513, a 43 ft. high centrifugal discharge unit. This elevator feeds directly to the calciner R-501.

The calciner is a refractory lined, oil fired, rotary kiln, 7 ft. 6 in. ID and 120 ft. long. Rotational speed is variable between 1.5 and 2 RPM using a variable diameter pulley drive, and the kiln has a slope of 3/8 in. per foot. The calciner product empties into four tube coolers attached to the shell and equipped with internal flights to contact the existing hot regenerated magnesia with incoming air which serves a secondary combustion air in the calciner, thus cooling the calciner product before it empties to the MgO conveyor #1, O-503.

This is a 9 in. diameter helicoid screw conveyor, 54½ ft. long, which elevates the product from the calciner tube cooler discharge seventeen feet to an enclosure housing the MgO processing equipment. The regenerated MgO is discharged from O-503 to a magnetic pulley for tramp iron separation. The MgO then passes through a 1 in. mesh vibrating screen and enters a high speed pulverizer which reduces it to the finished grind shown in Table 5.

The cleaned, cooled, and pulverized regenerated magnesia flows by gravity to the MgO elevator O-504, a continuous discharge design, 89 ft. high, which brings the material to the top of the MgO storage silo where it is loaded by means of a horizontal screw conveyor (MgO Conveyor #2) O-505 equipped with a 9 in. helicoid screw.

The MgO silo, I-502, 25 ft. in diameter and 45 ft. high is equipped with vibrating hopper bottom, O-506, and elevated on a structural steel support for direct gravity loading of the returning trucks.

The gas from the calciner, containing SO_2 and products of combustion, as well as a small percentage of excess air, is first partially cleaned of particles in the cyclone dust collectors P-075. This is a dual cyclone array, designed for a 1 in. pressure drop. The collected solids are returned to the calciner with the feed to the unit.

The partially cleaned calciner gas containing 8-10% SO_2 is further cleaned in a venturi scrubber of Chemico's special design, operated at a pressure drop of approximately 25 in. of H_2O where it is also adiabatically saturated. Next the gas enters the separator tower section, which is an integral part of the venturi equipment. The lower section of this 4½ ft. diameter vessel serves a cyclonic liquid separator and the upper section, containing eight feet of 3½ in pall rings is irrigated with cooled weak acid to further reduce the temperature of the gas to $100^{\circ}F$ in order to maintain the acid plant's water balance. A slip stream of cooling liquor is stripped of dissolved SO_2 in the weak acid

TABLE 5

DRY SCREEN ANALYSIS

Screen Size (Tyler)	Unpulverized Calciner Product	Pulverized Calciner Product
+50	30.3 %	5.3 %
+100	18.1	9.5
+200	9.4	8.8
+325	13.1	49.8
-325	29.1	26.6

stripping tower P-502, a small (15 in. diameter x 14 ft. high) packed contactor.

The stripped SO_2 joins the main gas stream and is ducted to the acid plant in 18 in. diameter, FRP pipe.

3.3 ACID PLANT

The regeneration system was installed at the Rumford acid plant of Essex Chemical Company located in Rumford, RI. The Rumford plant is located 55 miles from the installation at Mystic Station and haulage of the dryer and calciner products between these two sites was done by truck. Both plants were equipped to load their respective products from elevated silos and receive their feed by pneumatic unloading.

The plant has been producing sulfuric acid since 1929 when it was built by Chemical Construction Corporation and for the past several years all acid produced was sold in the merchant market with no captive use. Markets for the acid made at this plant are manufactured of detergents, dyestuffs, pharmaceuticals, aluminum sulfate, tanning chemical, steel pickling, boiler water treatment, lead-acid batteries, galvanization, etc. Sales in this market were at the full published price, at the time of this program, approximately \$46 per ton based on 100% H_2SO_4. No change in use or pricing was imposed when marketing the acid produced from MgO regeneration.

The plant is a contact sulfuric acid plant which uses sulfur as raw material. In the Chemico process the molten sulfur is injected into the furnace using a spray burner, with the process air first dried in a tower against 93% sulfuric acid. The catalyst is vanadium pentoxide.

When first built in 1928 the Rumford plant was an indoor plant typical of its time and had a capacity of about

20 tons per day of 100% sulfuric acid. Substantial modifications were made to the plant in 1948 by Chemico to increase its capacity to 50 tons per day. The modifications included improved converters, a converter heat exchanger, waste heat boiler and economizer all installed outdoors. In addition, the cast iron cooling section was enlarged and moved outside.

This was the plant that was modified in 1971 to accept the calciner off-gas essentially converting it to a metallurgical (roaster) gas plant; however, in the modification the capability to continue to burn sulfur and augment the SO_2 from the regeneration plant was retained.

Typical feed gas analyses to the acid unit are shown in Table 6.

The regeneration of magnesia, described in the previous section, produces an off-gas from the calciner approximately 100 fold richer in SO_2 than the power plant combustion gases treated in the SO_2 Absorption System. The calciner gas is of sufficient strength to be used as a feed for the manufacture of sulfuric acid.

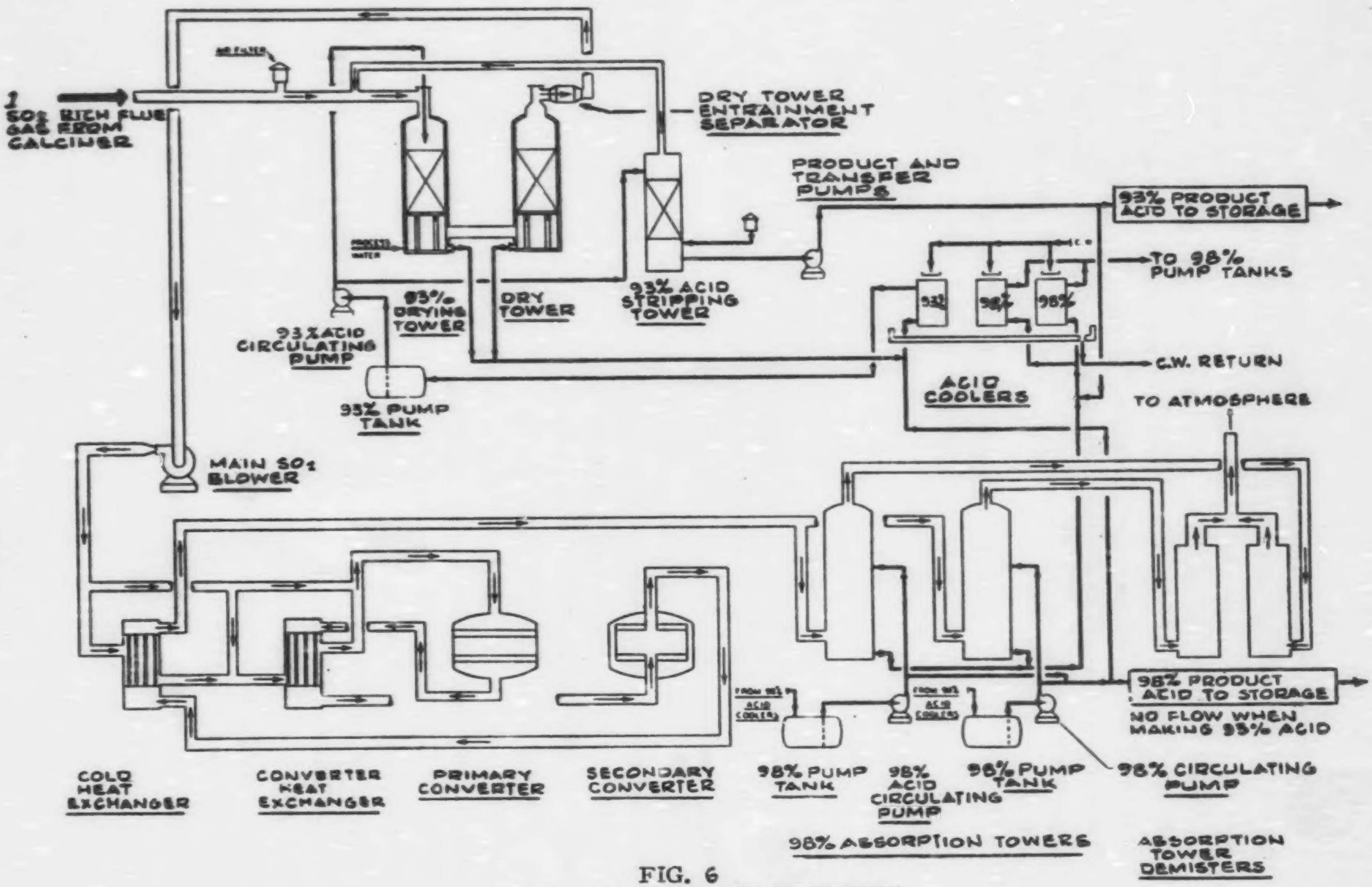
The small, sulfur burning acid plant (Fig. 6) required some modifications to enable it to accept the calciner gas. Provisions were also made during these modifications allowing the plant to burn sulfur as an alternative source of SO_2 or to operate on a combined feed from both combustion of sulfur and gas from the regeneration plant.

3.3.1 Acid Plant Modifications

The principal element replaced in the acid plant was Main SO_2 Blower, K-901. The original blower handled only air required for the conversion of SO_2 to SO_3 and was not designed to be gas tight (as required when feeding the acid plant with gas containing SO_2), or capable of the required

TABLE 6
SULFURIC ACID UNIT FEED GAS COMPOSITION, MOLE %

Source Of Gas	N_2	CO_2	O_2	H_2O	SO_2
Regeneration Section	73	6	5	7	9
Sulfur Burning	79		12		9



SO2 RICH FLUE GAS FROM CALCINER
AIR FILTER
PROCESS WATER
DRY TOWER ENTRAINMENT SEPARATOR
PRODUCT AND TRANSFER PUMPS
93% PRODUCT ACID TO STORAGE
TO 98% PUMP TANKS
C.W.
93%
98%
98%
C.W. RETURN
ACID COOLERS
TO ATMOSPHERE
93% DRYING TOWER
DRY TOWER
93% ACID STRIPPING TOWER
93% ACID CIRCULATING PUMP
93% PUMP TANK
MAIN SO2 BLOWER
FROM 98% ACID COOLERS
FROM 98% ACID COOLERS
98% PRODUCT ACID TO STORAGE
NO FLOW WHEN MAKING 93% ACID
COLD HEAT EXCHANGER
CONVERTER HEAT EXCHANGER
PRIMARY CONVERTER
SECONDARY CONVERTER
98% PUMP TANK
98% ACID CIRCULATING PUMP
98% PUMP TANK
98% CIRCULATING PUMP
98% ABSORPTION TOWERS
ABSORPTION TOWER DEMISTERS

FIG. 6

SULFURIC ACID PLANT — PROCESS FLOW DIAGRAM
AT ESSEX CHEMICAL CO RUMFORD, R.I.

control of suction pressure for operation of the regeneration plant. The replacement blower is an axial flow compressor of 5,240 CFM capacity driven by a 200 HP motor. It is capable of a suction pressure of up to 44 in. H_2O for the venturi pressure drop, duct and equipment losses, and calciner draft at the regeneration plant, and a 75 in. H_2O discharge pressure for the acid plant. The unit was designed to handle either air or a typical feed gas as shown below:

ACID PLANT FEED GAS	
SO_2	6.8%
O_2	9.4%
N_2	74.7%
CO_2	8.1%
CO	0.9%

The blower was equipped with an adjustable recycle control in order to accommodate the variable feed rate of the regeneration plant.

Another important element added to the acid plant was the Cold Heat Exchanger E-901. This piece of equipment supplies the heat to the incoming cold calciner gas equivalent to that available in the gas when burning sulfur in the sulfur furnace. This is necessary to ensure that the gas entering the first mass of the primary converter is hot enough to sustain the reaction. The exchanger, E-901, is a vertical shell and tube unit 3 ft. diameter x 9 ft. high containing 1700 sq. ft. of surface. Hot gas exiting the secondary converter enters the tube side on E-901 and heats the calciner gas passing through the shell side to 540°F.

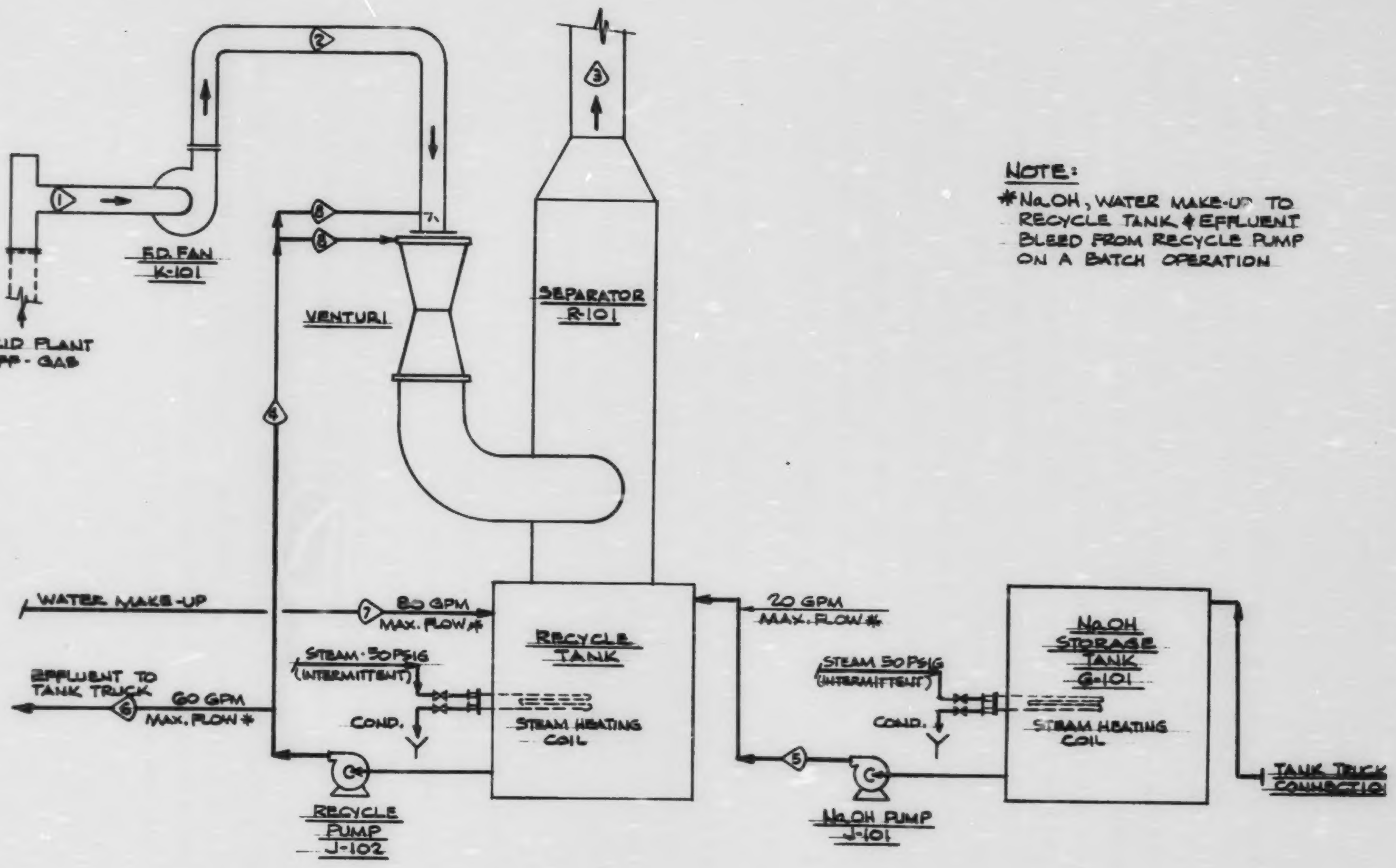
3.3.2 Acid Production

In the acid plant, calciner gas enters (Stream 1 on Fig. 6) and is diluted with sufficient air through the air filter for conversion and passes to the acid plant's drying tower where it is contacted with 93% sulfuric acid to remove the water from the gas. Entrained liquid is removed in F-901A, the Dry Tower. Any SO_2 absorbed by the 93% acid is stripped in the new Tower (F-903), a 30 in. diameter x 15 ft. high ceramic tower containing a 10 ft. bed of 1½ in. saddles, and returned to the main gas stream. The small amount of additional air is used for trim of the oxygen concentration. The gas is pressurized by the SO_2 blower and enters the Cold Heat Exchanger, passes to the Converter Heat Exchanger, E-902 and finally enters the converters H-901 A & B. The heat balance for the conversion step is maintained by using the gas from the second mass of the Primary Converter, H-901A, to heat the gas from the Cold Exchanger, and the gas leaving the fourth mass of the secondary converter, to heat the incoming feed.

Next, the stream containing 6.7% SO_3 is contacted with 98% H_2SO_4 in the Absorption Tower F-902 A & B. The gas streams from these towers are first demisted in R-902 A & B, then the remaining SO_2 is removed in the Tail Gas Scrubbing section shown in Fig. 7 and Table 7 to reduce the concentration of SO_2 leaving the plant to allowable levels.

The absorbant used in the Tail Gas Scrubbing section is NaOH solution.

In order to allow a rapid change to sulfur burning, when the plant is operated on 100% calciner gas, the sulfur furnace was equipped for oil firing. The combustion products from this operation are vented to the atmosphere through a short stack which could be bypassed. In normal operation, however, sulfur was burned concurrently because of a low $MgSO_3$ feed rate. This procedure had the advantage of allowing a rapid change should the calciner gas flow be interrupted. This alternative was possible because of the several dampers which had been incorporated in the ductwork.


1
2
3
4
5
6
7
8
8
ACID PLANT OFF-GAS
F.D. FAN K-101
VENTURI
SEPARATOR R-101
WATER MAKE-UP
80 GPM MAX. FLOW #
EFFLUENT TO TANK TRUCK
60 GPM MAX. FLOW #
STEAM-50 PSIG (INTERMITTENT)
COND.
RECYCLE TANK
STEAM HEATING COIL
RECYCLE PUMP J-102
20 GPM MAX. FLOW #
STEAM 50 PSIG (INTERMITTENT)
COND.
NaOH STORAGE TANK G-101
STEAM HEATING COIL
NaOH PUMP J-101
TANK TRUCK CONNECTION

NOTE:
NaOH, WATER MAKE-UP TO RECYCLE TANK & EFFLUENT BLEED FROM RECYCLE PUMP ON A BATCH OPERATION

FIGURE 7

TAIL GAS SCRUBBING SYSTEM

TABLE 7
STREAM PROPERTIES AND COMPOSITION
TAIL GAS SCRUBBER SYSTEM

STREAM NUMBERS		1	2	3
Temperature	°F	160	165	77
Pressure	IN WC	0	14	0
Flow	ACFM	4,250	4140	5600
Dry Gas	lb/Min	279	279	276.3
H_2O	lb/Min	0	0	5.6
Wet Gas	lb/Min	279	279	281.9
SO_2 (Design)	lb/Min	3.0	3.0	0.3
SO_2	PPM Dry By Volume	5000	5000	500

STREAM NUMBERS		4	5	6	7	8
Temperature	°F	77	Amb.	77	Amb.	77
Flow	GPM	152	0.6	3.0	3.0	76
NaOH	lb/Min		3.75			
Na_2SO_3	lb/Min	299		5.9		149.5
H_2O Liquid	lb/Min	1266	3.75	23.6	24.6	133
Total Solution	lb/Min	1565	7.5	29.5	24.6	782.5

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4.0 POLLUTION ABATEMENT SYSTEM PERFORMANCE

Construction and check out of the system at Dickerson was completed by September of 1973 and gas was first passed to the unit on September 14th. In the following three month period 600 hours of operation were logged in the initial operation and debugging of the system. Most operating periods were of short duration, from 10 to 24 hours, before a system problem would force a halt. The major problem area encountered in the earliest runs was the MgO feed system, where plugging occurred in both the mix tank and slurry lines. The plugging was partially remedied by two modifications: the installation of a premix tank before the slaking tank to assure complete wetting of the MgO powder with recycle liquor, and addition of a steam sparging system to the tank. Both these remedies have also been found necessary to correct similar problems in the oil-fired application.

Other shut-downs in this period were caused by leaks that developed in the venturi scrubber internal piping between first and second stages. The leaks occurred in expansion joints where sections of the recycle lines left the first stage reservoir. Testing showed the joints supplied were off-specification stainless steel that was not resistant to this acidic stream.

Despite these problems two sequential six-day runs interrupted by a 24 hour boiler outage, were attained in the preliminary operations. While these runs did show the operability of the entire system their usefulness in assessing chemical stability was limited because of high entrainment losses from the venturi which prevented equilibrium conditions from being attained in the recycle streams. A low SO_2 removal efficiency was also noted in the preliminary testing.

An inspection of the scrubber in January 1974 showed that off-specification material, unresistant to the acidic environment of the first stage, had also been used in bolts, nuts, hanger rods, spray nozzles and some piping in the vessel. A major shut-down extending over the next six months was taken to replace corroded parts and repair damaged areas. A policy was also adopted to test each replacement item to ensure that it was of the specified type 316 stainless steel. A brief verification run was made at the beginning of May and the new problems which it uncovered were corrected in the following weeks, prior to starting regular operations.

During this initial operating period MgO regeneration was not available to the Dickerson plant. The EPA unit located at Rumford, Rhode Island was being used exclusively for regeneration of the MgO from the oil-fired system prototype. The regeneration plant was not scheduled to be available for Dickerson Station till July 1, 1974, when PEPCO would then have exclusive use of the facility. In anticipation of this use, PEPCO sent three rail cars of $MgSO_3$ from its operations to the Rumford facility for preliminary testing of magnesium sulfite derived from coal fired operation.

Before operations with regenerated MgO were started at the power plant some additional test work was conducted in June 1974. The low SO_2 removal efficiency, first noted in the debugging runs, was confirmed by these test runs and attributed to a low (1" to 3" ΔP) second stage pressure drop. A restrictor, to reduce the second stage throat area by 40%, was designed in order to increase the pressure drop to design value.

The test runs exhausted the initial supply of MgO which had been delivered for start up operations. The abatement system was idle for another six week period, awaiting a new shipment of MgO, the return of regenerated magnesia from Rhode Island and installation of the throat restrictor.

4.1 DEVELOPMENT TEST PROGRAM

The silos at the regeneration plant were completely emptied of all Boston Edison $MgSO_3$ in July 1974 and filled with PEPCO material from the rail cars. (The lack of any size reduction or lump breaking equipment in the path between the Dickerson dryer and regeneration plant calciner caused unloading problems then, and throughout the entire program). The operation, after a brief one-day adjustment period, produced a calciner product which tests showed to be of acceptable quality for recycle.

When the PEPCO $MgSO_3$ was processed in the calcining facility, it was found necessary to add carbon to the feed to effect the reduction of magnesium sulfate. The oil derived calciner feed had contained sufficient carbon to satisfy the reduction reaction needs as the absorber, in the oil-fired application, treated the gas directly from the boiler. This "scrubbed" out the soot and uncombusted fuel associated with oil-fired applications which not only reduced the power plants particle emissions but also provided the 0.5 to 1% carbon required in the process for reduction of $MgSO_4$.

Because of the necessity to prevent fly ash contamination of the recycled magnesia in the coal-fired application, the incoming flue gas at PEPCO was precleaned (by ESP and/or scrubber) to remove the fly ash. This also removed any carbon sources from the flue gas before it entered the second stage.

A supply of petroleum coke was located and added in the required proportions with the calciner feed. This was continued throughout the integrated operations at the acid plant with Dickerson Station.

These initial operations at the regeneration plant also saw the return of the "slide" problem which had been encountered and partially solved for the oil-fired application. (A "slide" was a description of a rapid emptying of a part of the calciner charge before decomposition of all the $MgSO_3$ in it). The slides caused overloading of the materials handling equipment on the kiln discharge, when they occurred, which forced a reduction in the kiln feed rate. Analysis of the product after a slide showed several percent of undecomposed $MgSO_3$. The frequency of occurrence of slides was eventually related to the size distribution of the entering feed. It was shown that a high percentage of fines in the feed triggered them. By adjusting the methods of operation of the dryer to reduce the amount of fines in the dried $MgSO_3$ during the oil-fired application calciner feed rates up to 80 lb/min were able to be achieved.

A different type dryer than the one used in the oil-fired boiler application had been designed for the PEPCO operation. (A counter-current design dryer had been installed at Boston Edison Company's Mystic Station while a co-current design dryer was installed at Dickerson Station). This difference in dryer types was one of the major equipment differences between the oil-fired boiler and coal-fired boiler applications in the MgO slurry dewatering process.

Analysis of the product from the PEPCO dryer showed that a higher percentage of fines were produced from the co-current dryer and that the fines could not be controlled by the techniques used at Mystic Station. The slide problem limited calciner feed rates to 40 lb/min rather than the 106 lb/min design rate. Because of the interrupted operations of the plant a dryer investigation could not be undertaken during the program, and calciner feed rates were held at 40 lb/min. However, this rate was sufficient to keep the FGD operation supplied with recycle MgO.

At Dickerson Station a shipment of virgin MgO was received July 29th and the first regenerated MgO was returned August 16th. With this inventory of material the performance testing phase was started to provide preliminary reliability data and optimization information.

Tests conducted over the following two weeks checked SO_2 removal efficiency prior to installation of the throat restrictor and tested a new configuration for MgO pre-mixing using an eductor mixer. (The mixing test was inconclusive because the unit obtained was too small and plugged). The MgO slurrying system was returned to the original configuration, an agitated pre-mix tank. The throat restrictor was also installed in preparation for the series of performance tests which were to follow.

The first six series of performance tests were started during the last week of August and the test program continued through the months of September with 400 hours of operation logged. Test series conditions included full to half boiler load with full to half design flow to the scrubber/absorber, and inlet gas taken after the precipitators or with no pre-cleaning Series designations are shown in Table 8 . A description of these tests and the results are presented in another section of this report. Also incorporated into the foregoing investigation was an analysis program of the first stage recycle liquor, thickener and pond and river discharge flows

TABLE 8

PERFORMANCE TEST DESIGNATIONS

Test Series	Boiler Load (% of Design)	Scrubber Gas (% of Design)	Scrubbed Gas Pretreated
5A	100	100	E.S.P.
5B	50	50	E.S.P.
6	100	50	E.S.P.
7	100	100	NONE
8	100	50	NONE

for a number of chemical species of environmental significance.

Following these tests the plant was readied for the balance of the optimization program. Emphasis was placed on the assessment of the effect of coal fly ash on the regeneration and utilization of the recycled magnesia. A regular sampling and analysis program was initiated for flue gas monitoring and chemical determination of all major stream compositions. Over the next four months, system operations totaling 1350 hours provided much information on the effect of regeneration operations on the absorption of sulfur dioxide, operation of the centrifuge and dryer and the conditions of scrubber/absorber operation.

Attainment of the major goals of the project, multiple recycles of magnesia and conclusive demonstration of system reliability were thwarted by the deteriorated condition of the piping which evidenced itself almost immediately after entering this phase of the demonstration program. As the performance testing was completed the first of many failures in the rubber lined, first stage recycle piping occurred forcing a shut-down of the system. During the rest of the operations first and second stage pipe failure was the most common shut-down cause. A summary of operations during this period is given in Table 9 .

Some improvements were obtained when several sections of the second stage recycle pipe were replaced but leaks in the other sections and the unrectified first stage problem limited operating periods to days rather than weeks. The longest continuous run during this period was eleven days from December 13th through December 23rd. Operations resumed on December 27th and continued into January 1975 with tests of first stage operation alone. These tests also

TABLE 9

SUMMARY OF OPERATIONS FGD SYSTEM PERFORMANCE TEST PHASE

Week Ending	Goal	Problem Areas	Number Of Interruptions	Avg. Load	Hours Of Scrubber Operation
11/01	System Repair				0
11/08	Repair				0
11/15	Operate at design load	Centrate Pump	1	60-80%	88
11/22	Establish dependable & continuous operation	Bucket Elevator(1) & Piping Leaks	2	60-70%	84
11/29	Continuous Operation	Dryer Fan Drive	1	70-80%	123
12/06	Continuous Operation	Piping Leaks	1	75-80%	77½
12/13	Continuous Operation	Piping Leaks & valve failure	2	60-80%	114½
12/20	Continuous Operation	Piping Leaks	1	60-85%	159½
12/27	Continuous Operation	Piping leaks & valve failure	2	60-90%	83½
01/03	Continuous Operation	Piping leaks	1	80-90%	53
01/10	Operate 1st Stage only			90-100%	78
01/17	Continuous Operation	MgO inventory depleted	1	60-70%	75
01/24	Continuous Operation	Piping Leaks & Conveyor Jam	1	50-60%	30
01/31	Continous Operation	Piping Leaks	1	50-60%	37½
			OVERALL AVAILABILITY 46.1%		1003.5

1) Marginal bucket elevator capacity limited operation to average 70% of design

provided data on first stage SO_2 removal capacity. A final series tested recycle liquor flow effect on SO_2 removal efficiency.

Operation of the regeneration plant during this period was sporadic because of the outages at Dickerson and the resultant interruptions in feed to be calcined. The major mechanical problem encountered at the regeneration plant was a failure of the pinion gear in the main drive on the calciner (due to misalignment and lack of lubrication). A replacement was locally fabricated and the unit was only out of service for two weeks during December. A summary of operations at the regeneration facilities is given in Table 10.

During the program approximately 2247 tons of $MgSO_3$ were processed to MgO at the acid plant.

During January 1975 plans were made for major overhaul of the PEPCO FGD system including replacement of all "thin" second stage pipe and repair of several sections of first stage rubber lined pipe. This work was scheduled in conjunction with PEPCO's planned 12 week overhaul of its No. 3 boiler and turbine. Repair of the FGD system was for the most part, completed by April. PEPCO, however, discovered a major defect in the No. 3 turbine during their boiler overhaul which necessitated a number of extensions of their boiler outage. Eventually the outage extended to the end of July 1975. The outage extensions created secondary problems in the financing of the project and the program was terminated.

TABLE 10

SUMMARY OF OPERATIONS CALCINER SYSTEM PERFORMANCE TEST PHASE

Week Ending		Operated For Hours
11/01/74	Maintenance on Bucket Elevator	0
11/08/74	Routine Maintenance	0
11/15/74	Routine Maintenance - Accumulate $MgSO_3$ Inventory	0
11/22/74	Routine Maintenance - Accumulate $MgSO_3$ Inventory	58
	START-UP	
11/29/74	MgO Production for PEPCO	47
12/06/74	MgO Production for PEPCO	87.4
12/13/74	MgO Production for PEPCO	24
12/20/74	Maintenance on Calciner Main Drive and MgO Production for PEPCO	8.9
12/27/74	Complete Repairs on Calciner Main Drive and MgO Production for PEOCO	0
01/10/75	Run Test on Boston Edison $MgSO_3$ without Coke Addition	100.9
01/17/75	Test PEPCO $MgSO_3$ without Coke Addition	44.2
01/24/75	MgO Production for PEPCO	4.5
01/31/75	MgO Production for PEPCO	7.9

4.1.1 October 1974 - Details

Regular operation of the FGD system at Dickerson was resumed after the performance test work on September 30. This first operation was stopped after three days when problems with the MgO premix tank forced an outage. The system was restarted at 1 p.m. on the 3rd and ran continuously for the next 12 days at approximately 50 to 80% of design load. A failure of a 14" rubber lined pipe elbow in the first stage recycle loop forced another outage at 4 p.m. on the 14th.

The analytical program was fully implemented by October 16, 1974. The scrubber ran from a start up at 12:00 a.m., October 16 to 9:00 a.m., October 18. A planned shutdown was taken to repair several leaks in the first stage recycle lines and to attempt to increase the flows in MgO slurry feed system. The scheduled repair was completed by noon, October 18, however, after operations resumed low flow of MgO slurry resulted in loss of pH control in the second stage. Leaks in the MgO magnetic flow meters also caused delays. Flue gas treatment resumed at 3 a.m. October 23, 1974.

During the night of the 24th problems again occurred in the MgO slurry system with line plugging reducing pH in the second stage to 6.6; however, the situation was corrected and the system ran till 9 a.m. on the 25th when the unit was shut down for a boiler tube leak. The boiler was repaired by October 28, but the scrubber system was waiting for maintenance and was not returned to service.

4.1.2 November 1974 - Details

During the 30 day period, November 1 to November 30, 1974, the scrubber operated 318.9 hours out of a possible 720 hours for a 44.3% availability. During the period of operation the unit ran at 50% to 100% of design gas load resulting in a total average available capacity for the month of November of 34%. Because of scheduled outage in the earlier portion of the month for leak repair and overall maintenance, most running time was logged in the latter portion of the month.

Scrubber Operation and Comments

The scrubber was brought on-line at 2 a.m., November 10, at 80% of design flow. Attempts to increase the gas flow to full load were hindered, as problems in the MgO-slurry feed system prevented pH control at increased loads. The system operated steadily, adjusting the scrubber gas flow to maintain a pH of 5.7, till 2:20 a.m. on the 11th, when the scrubber was taken off-line due to a leak at the inlet of the second stage bleed line.

Repairs were completed at 8:00 p.m. on the 13th. The system was returned to operation and the scrubber load was varied between 60 and 80% to compensate for station load changes during this period. On November 15, at 10 a.m., a reduction to 1/2 load was made to repair a leak in the centrifuge. When the centrifuge was diverted, the solids in the second stage recycle slurry reached a high level and the low load was maintained until the solids level was centrifuged down. At this time it was noted that the flow to the centrifuge had dropped to 55 gallons per minute, which prevented operating the unit at design gas loading. This was corrected by operating with both the main and spare pump. The plant continued in operation with the scrubber load gradually increased to 100%, holding pH in the 7.0 to 7.3 range. (Normal operating practice was to control pH by maintaining a steady MgO feed, adjusting the gas flow to maintain the pH set point affected by variations in the coals sulfur content).

A leak developed in a first stage recycle control valve and the unit was taken off the line to make repairs on the 16th. Operations resumed on the 19th and continued through the 21st when the belt drive failed on the dryer I.D. fan.

The fan repairs were completed and gas was put back into the scrubber at 3:15 p.m. on November 23, but by 11 p.m. the No. 3 boiler developed a tube leak, forcing a shutdown. This was repaired and operations resumed by 2:40 a.m. on the 24th, six hours later another interruption was needed to repair a bearing on the dryer product screw conveyor. The system was returned to operation 2:45 that afternoon and operated continuously through the end of the month. Operating adjustment had to be made during this period to accommodate generating station coal feeder problems.

During this final fifteen day period the scrubber operated a total of 241.6 hours out of a possible 360 hours for a 67.1% availability (51.7% average available capacity based on a 77% of design gas flow).

4.1.3 December 1974 - Details

During the 31 day period, December 1, 1974 to December 31, 1974, the scrubber operated 430.8 hours of a possible 744 hours, an availability of 57.9%.

During the month of December various new studies were initiated and the wet chemical method for the specific determination of $MgSO_3.6H_2O$ and $MgSO_3.3H_2O$ in the mixture was used to determine the hydrate ratio in the centrifuge cake samples. Previously, the ratio of $MgSO_3.6H_2O/MgSO_3.3H_2O$ had been approximately determined on the basis of percent combined water (water of crystallization) as determined on an OHAUS moisture balance.

Chloride in the system was also measured as a function of operating time.

Scrubber/Absorber Operation and Comments

The run which had started on November 24 had to be aborted on December 1 after a leak developed in the rubber-lined bleed line to the thickener area. Operations were again resumed on December 5 and continued till 4:30 a.m. on the 11th when leaks in the first stage bleed line again forced an outage. During this period adjustments in operating conditions were used to control problems of MgO feed, station coal handling and centrifuge cake buildup in the discharge hopper.

Operations were resumed at 6 p.m. on the 13th and this run continued till the 23rd when leaks in both the first and second stage discharge headers forced a shutdown.

The system was restarted on the 27th but the run was aborted on the 29th after 64 hours of operation with another recycle pipe failure.

4.1.4 January 1975 - Details

Measurements of the pipe wall thickness and observations of the pipe involved in the numerous failures which had been encountered showed that a major system overhaul and pipe replacement program was necessary if any sustained operations were to be achieved. PEPCO has rescheduled an overhaul of the No. 3 generating unit to 1975. This was a five year inspection and overhaul of the generator originally due in 1973. Plans were made to recondition the scrubber system during the boiler outage scheduled to start in January 1975. Some additional test work was conducted with the pollution abatement system during January while waiting for the planned boiler shutdown.

These tests included four days of operation with first stage (particle control) only, and another series of tests to determine the effect of variation of recycle flow rate on SO_2 removal efficiency.

4.2 UTILITIES CONSUMPTION

Electric power consumption (in MW/Day) of the particle control and flue gas desulfurization system installed at Dickerson Station is given in Table 11 for the major operating periods of the test program.

Also included in the table are data on average fuel consumption for the dryer used in the FGD system along with an approximate gas flow processed during the period. While there is some effect caused by the variation in fuel sulfur content on the amount of centrifuge cake produced it appears that a significant factor in dryer fuel oil consumption is the weather. Fuel consumption is seen to increase during the winter months and decrease during the summer months while the gas flow processed through the system remained the same.

4.3 MgO CONSUMPTION

Magnesia consumption during the fifteen month period of operation of the plant was high. It is evident that the loss rate during the period of initial operations was higher than the loss rate during the latter planned operations phase.

MgO shipments to the plant were made corresponding to these two periods. An initial inventory of 357 tons of MgO was shipped to Dickerson between August and December of 1973 for the start-up operation. An additional 281.3 tons of virgin magnesia wereadded as make-up between July and November 1974 for the planned operations phase. The addition of make-up MgO at Dickerson between July and November is given in Table 12.

TABLE 11
UTILITIES CONSUMPTION
FGD SYSTEM

Major Operating Periods	Approximate Gas Flow % of Design	Average Electric Power Consumption MWH/Day	Average Dryer Fuel Consumption GPM
April 25-30, 1974		38.6	0.57
June 13-16, 1974		39.8	0.39
Aug. 13-17, 1974		34.8	0.33
Sept. 3-11, 1974		27.7	0.38
Oct. 16-18, 1974	66%	41.8	0.72
Nov. 24 - Dec. 1, 1974	75%	53.3	0.93
Dec. 6-10, 1974	70%	49.0	0.95
Dec. 14-23, 1974	67%	45.5	0.91
Dec. 27-29, 1974	80%	57.5	0.96
Jan. 12-14, 1975	70%	50.0	0.93
Aug. 12-14, 1975	70%	50.0	0.54
Aug. 22-26, 1975		33.8	0.32

TABLE 12
MgO MAKE UP
FGD SYSTEM

Date	Quantity Of Virgin MgO Received
7/29/74	67.6 Tons
9/1/74	73.4 Tons
10/24/74	67.4 Tons
11/4/74	72.9 Tons
	281.3 Tons

4.3.1 Losses - Preliminary Operations

Estimation of dryer product generated at Dickerson Station based on shipping records showed 485 tons of $MgSO_3$ were produced from the initial charge of 357 tons of MgO. This accounts for only 55% of the dryer product that should have been produced based on the design material balance.

An inventory evaluation also showed that 144 tons of MgO equivalent were lost during the same period giving a loss rate of 40.4% of the material fed. Since no magnesia regeneration was being performed during the early period the losses were essentially confined to the FGD system site. Part of the loss can be attributed to a number of unmeasured spills and discharges during this period, some of these were:

1) Material discarded when cleaning out the MgO slurry system after a plug.
2) Material lost when draining the scrubber for repair work.
3) Spills of dryer product due to material handling equipment overloads.

Another major loss point was entrainment and overflow from the second stage of the scrubber. The few chemical analyses of recirculating slurry taken during the early period showed a maximum of 8% magnesium sulfate in this stream. Since a concentration of 15% to 18% $MgSO_4$ is expected in the equilibrium system the low concentration indicated a large bleed (or leak) from the magnesia recycle loop.

Chemical analysis of the fan sump discharge during the performance testing conducted in August and September 1974 showed high levels of magnesium salts and solids in this stream. Analysis showed $MgSO_4$ concentrations as high as

10% again indicative of excessive entrainment. The entrainment was probably the major loss point sustained during these preliminary operations.

4.3.2 Losses - Planned Operational Testing Period

Corrective action was taken to reduce the entrainment from the scrubber system by lowering the operating liquid level in the second stage reservoir from 80% to 40% to reduce gas velocity in that zone. Analysis of the fan sump pit discharge in early October 1973 showed much lower $MgSO_4$ concentration. Losses from this source were reduced to less than 0.5% (based on Mg^{++} content of the fan discharge) during the final three months.

4.4 MgO REGENERATION

MgO inventory recording was started at Dickerson in October 1974 and the weekly inventory records are summarized for the period of operation in Tables13 and14 .

Test work at Dickerson in August had been carried out using regenerated MgO; however, the losses, which had occurred before the development program was initiated, necessitated replenishment of the inventory with virgin MgO. During September and October 141 tons of MgO were received and 109 tons of regenerated alkali were returned from Rumford for operation of the integrated system. During October the Dickerson system operated 420 hours of a possible 647 hours, consuming 80% of the alkali available and producing 482 tons of dryer product.

Thus, at the start of November, 141.4 tons of MgO and 100.6 tons of $MgSO_3$ were on hand at Dickerson. Of the MgO, approximately 65% had been regenerated at least twice. Before operations were resumed at Dickerson 73 additional tons of virgin MgO was received and added to the silo. Inventory on November 8th, for the resumption of Phase II of the development program, was 213.3 tons of MgO and 100.6 tons of

TABLE 13

INVENTORY SUMMARY BY WEEK-FGD SYSTEM

	$MgSO_3$ (Tons)		Closing (Tons)	
Week Ending	Produced	Shipped	MgO	$MgSO_3$
11/1/74	0	19.5	110.4	56.6
11/8/74	0	0	182.3	56.5
11/15/74	32.8	20.9	150.5	68.4
11/22/74	98.4	35.4	105.2	131.4
11/29/74	142.5	141.2	110.1	123.3
12/6/74	54.0	141.4	131.0	14.0
12/13/74	116.8	35.1	68.2	95.9
12/20/74	157.9	108.6	26.9	145.1
12/27/74	83.6	114.6	73.9	96.2
1/3/75	65.0	0	48.9	161.2
1/10/75	0	0	48.9	161.2
1/17/75	85.9	88.5	54.8	158.5
1/24/75	8.0	76.3	118.3	90.2
1/31/75	---	43.2	105.1	71.5
		824.7		

TABLE 14

INVENTORY SUMMARY BY WEEK-REGENERATION SYSTEM

Week Ending	MgO (Tons) Produced	MgO (Tons) Shipped	Closing (Tons) MgO	Closing (Tons) $MgSO_3$
11/1/74	0	11.9	31	44
11/8/74	0	0	31	44
11/15/73	0	0	31	44
11/22/74	30	0	61	8
11/24/74	26.4	91.7	11.5	45.1
12/6/74	50.3	35.5	26.3	52.6
12/13/74	13.3	27.8	11.8	62.1
12/20/74	4.9	15.7	0	156
12/27/74	0	0	0	233.4
1/3/75	48.2	12.8	35.4	126.4
1/10/75	56.5	0	90.9	0
1/17/75	30.6	53.8	67.7	7.6
1/24/75	2.5	71.0	1.5	85.0
1/31/75	4.4	0	5.9	135.5

$MgSO_3$ (total MgO equivalent 250.6 tons). The approximate inventory on December 27th, after 730 hours of operation, was 73.9 tons of MgO and 329.6 tons of $MgSO_3$ (total MgO equivalent 196 tons). The inventory difference of 54.6 tons is attributed to process losses, approximately 150 lb/hr. This may be contrasted with 223.7 lb/hr. loss sustained in the final operational period of the New England SO_2 Control Project. As was the case in the oil-fired application, most of these losses occurred from uncontrolled spills and bleed streams at the regeneration plant; however, during this period, November to January the regeneration plant only operated a total of 226.3 hours accounting for the lower loss rate during the PEPCO project.

By the end of December the entire inventory had been processed through an additional cycle. Some limited runs were conducted in January with 142 hours of two stage operation at Dickerson which initiated a new cycle of regeneration.

At Rumford, 194 tons of $MgSO_3$ remaining from the New England SO_2 Control Project had been in storage since July 1974. The first rail car of this material (80 tons) was processed with dryer product from Dickerson during January 1975 in a rate check of the calciner feed capacity.(1)

An examination of the statistical summaries for the Dickerson operations reveals a trend of a gradual increase in unreacted MgO in the recycle liquor similar to that seen in operations at Boston Edison's Mystic Station.

Period	MgO in Recycle (Avg.)	Standard Deviation
11/10-11/29	3.81%	2.90
12/1-12/29	3.95%	3.10
1/11-1/27	5.79%	1.40

4.5 POST OPERATION FGD SYSTEM INSPECTION

In January 1975, before the FGD system overhaul, the major pieces of equipment in the plant were inspected to determine their condition. The scrubber vessel was found to be in good condition with only a few, small internal areas where its plastic lining had failed. The fan was found to be in excellent condition as was the secondary mist eliminator vessel. Most other equipment, vessels and pumps in the system were in good to satisfactory condition. Major failures and unsatisfactory conditions were noted for the first and second stage recycle piping. A report of the inspection is given in Appendix 1.

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(1) Due to lack of operations at PEPCO and eventual termination of the entire program the regenerated MgO never found use at Dickerson Station.

An operating history of the plant showing the number of runs started and their duration is given in Table 15.

Problems which had an adverse effect on operability of the FGD system were divided into several catagories which are described in the following sections.

4.5.1 1st Stage Piping

Problems with some portion of the 1st stage piping system were the most frequent cause of scrubber shutdowns and accounted for about one-third of the total. There are five separate, but interrelated causes of failure that have been recognized, as follows: (a) pulsation in the piping near the inlet to the scrubber, (b) high liquid velocities at some points in the system, (c) the presence of sharp objects in the system due to corrosion of stainless steel, (d) rubber lining thinner than specified, (e) poor adhesion of rubber lining to the steel pipe.

(a) Pulsation in the piping near the inlet to the scrubber caused several failures of the pinch valve liners. The valves were finally replaced with rubber lined spool pieces. The pulsation also contributed to the failure of the elbows just before the scrubber inlet in all three of the 6-inch lines to the center nozzle. The cause

TABLE 15

OPERATING HISTORY
FGD SYSTEM

Date	Number Of Runs Started	Duration Of Runs, Hours
1973:		
Sept.	2	46
Oct.	2	90
Nov.	4	142
Dec.	3	249
1974:		
Jan.	1	192
Feb.	0	0
Mar.	0	0
Apr.	5	174
May	1	100
June	0	0
July	0	0
Aug.	6	248
Sept.	6	326
Oct.	4	344
Nov.	5	318
Dec.	6	420
1975:		
Jan.	3	131
TOTALS	48	2,780

of the pulsation was eventually traced to the hydraulic design of the piping system. The pressure drop between the high point in the system and the scrubber inlet was less than the vertical distance, resulting in an alternate filling and emptying of the line. Corrective action has been taken by the installation of restrictive orifices in each of the inlet pipes, immediately before entrance to the scrubber. Also, new tangential nozzles have been designed for a higher pressure drop--sufficient to compensate for the head between the restrictive orifice plate and the nozzle outlet.

(b) Excessive liquid velocities caused failures in the 1st stage bleed control valve and in a 316 stainless steel pipe section immediately after the control valve. The corrective action taken (sometime prior to August 1974) was the installation of three eccentric orifice plates in horizontal piping runs in order to increase the pressure at the downstream side of the control valve, thus permitting the valve to operate in a more nearly open position with reduced velocities through (and immediately after) the valve. The eccentric orifice plates were designed for velocities that would be adequately low for normal conditions. However, with the actual conditions, as will be discussed in (c), (d) and (e), failures of the rubber lining occurred adjacent to and downstream from the eccentric orifice plates. The plates were removed, leading to a recurrence of failures in and after the control valve. Corrective action has now been taken by the installation of a concentric orifice flange at the short, vertical, discharge end of the line.

(c) Sharp pieces of metal were present in the liquid cycle on numerous occasions as a result of corrosion failures of stainless steel as will be described in a subsequent part of this section. The adverse effects were aggravated by conditions to be discussed in (d) and (e).

(d) The rubber lining as supplied was thinner than the 1/4" specified. This was substantiated in December 1974 by micrometer measurements of rubber samples from five locations, selected at random from pipe sections that failed. Readings, in inches, were as follows:

1	2	3	Avg.
0.167	0.150	0.161	0.159
0.160	0.155	0.160	0.158
0.156	0.153	0.154	0.154
0.160	0.160	0.150	0.157
0.139	0.133	0.133	0.135
		Average	0.153

In every instance the original "waffle" pattern was still on the rubber and there was no evidence of any erosion. Replacement pipe linings have been a full 1/4-inch thickness and there have been no failures of any of the replacement linings.

(e) Adhesion of rubber lining to the steel has been poor. This has not been measured quantitatively but has been judged qualitatively by pulling portions of the lining from pieces that have failed and also by inserting a knife or fingernail between steel and rubber flanges. The significance of poor adhesion is that it allows a small failure to grow rapidly into a large one that cannot be controlled with external patches.

4.5.2 2nd Stage Piping

Failures of second stage piping became quite extensive in October 1974, and several replacements had to be made. Portions of the 16-inch and 20-inch piping that wore were repaired by applying a polyester or epoxy coating, reinforced with glass fabric to the thin area.

Failures are ascribed to a combination of erosion and corrosion. For long-term commercial use it is planned that rubber-lined piping will be specified.

An extensive series of piping thickness measurements were made in late December 1974, using an ultrasonic gauge, with the primary objective of establishing which portions of the piping system needed to be replaced (or covered) in order to provide for three months of additional operation.

4.5.3 Corrosion of Stainless Steel

During early operation of the scrubber there were a number of failures of stainless steel elements fabricated of type 304 or other types that are less corrosion resistant than type 316. Type 316 has performed with mixed results, depending on the specific environment encountered.

The first stage tangential inlet nozzles corroded and failed. It was concluded that the cause of failure in this case was the condensation of strong sulfuric acid from the flue gas onto the outer surface of the pipes--a mechanism similar to that which has been encountered in air heaters when flue gas is cooled to too low a temperature. Replacement nozzles were fabricated in both glass reinforced polyester plastic and stainless steel coated with glass reinforced polyester. Stress has been another cause of failure of type 316 stainless steel exposed to first stage liquid as evidenced by extensive attack where threads had been cut or cold bends had been made. This was particularly noticeable in the case of bolts in the south thickener and hanger rods supporting the first stage demister piping.

Replacements for the latter were heat treated after fabrication in order to prevent attack from this cause.

In the absence of either undue stress or conditions that permitted condensation of strong acid from flue gas, type 316 provided satisfactory performance. Examples are the hold-down straps for the mist eliminators, where there was general light pitting but no measurable loss of thickness and short sections of piping to the tangential nozzles, just before entrance to the scrubber, where thickness was very nearly equal to specifications and there was no visual evidence of corrosion.

5.0 PERFORMANCE TEST RESULTS

Performance tests of the particulate emissions and SO_2 control system installed at Dickerson Station were conducted by York Research Corporation. The testing period extended from August 26, 1974 through September 21, 1974 following a preliminary period in September for equipment set-up, familiarization, establishing gas velocity profiles and gas flow measurement.

The performance testing used methods detailed in the Federal Register, Vol. 36, No. 247, 1974 for:

a) Gas Analyses by Method 3
b) Particle Sampling Outlet by Method 5
c) SO_2 Removal Efficiency by Method 6
d) SO_3 Determination by Method 8

Particle sampling at the inlet was conducted in accordance with ASME PT6-27 (modified).

Sizing of the collected particle samples was done microscopically. This method is semi-quantitave as the actual deminsions measured are the diameter or length of the particle. This gives a general indication of the particles size and a mathematical approximation of its mass.

The data have been abstracted from York Research Corporation's Report Y-8513 dated January 31, 1975.

5.1 PARTICLE EMISSIONS

Results of the particle emissions tests are summarized in Table 16. Each result represents the average of four tests per series. Additional data are also provided on gas flow and composition.

TABLE 16

PARTICLE EMISSIONS TEST RESULTS

FGD SYSTEM

Test[1] Series	Boiler Load MW	ESP[2]	ΔP[3] in H_2O	Flue Gas Flow 10^3 ACFM		Temp. °F	CO_2 %	O_2 %	Gr/SCFD 12% CO_2	Gr/ACF
5A	183	Yes	10.2	299	Inlet	242	12.2	6.5	0.190	0.131
					Outlet	126			0.002	0.002
					Efficiency				99.0	98.5
5B	96	Yes	11.1	161	Inlet	241	11.9	6.7	0.101	0.070
					Outlet	116			0.005	0.004
					Efficiency				95.0	94.3
6	185	Yes	10.6	153	Inlet	240	12.3	6.4	0.120	0.085
					Outlet	136			0.007	0.007
					Efficiency				94.2	91.8
7	183	No	10.7	282	Inlet	245	11.9	6.5	3.654	2.474
					Outlet	118			0.014	0.011
					Efficiency				99.6	99.6
8	176	No	10.0	127	Inlet	232	11.6	7.1	3.276	2.212
					Outlet				0.017	0.014
					Efficiency				99.5	99.4

(1) Average of four (4) tests per test series

(2) Indicates whether the flue gas treated by the scrubber stage passed through the electrostatic precipitator (Yes) or not (No).

(3) First stage (scrubber) pressure drop.

As noted in Section 4.1 the conditions over which the tests were conducted were varied to provide data on a range of operational situations, taking gas either ahead of or after the precipitator.

These results show the low particle emissions attainable using the Chemico wet scrubber. In the tests taken after the precipitator outlet particle emissions were less than 0.01 gr/SCF for all cases despite the low inlet loading. For the full boiler load, full gas flow case (5A) fly ash emissions were less than 4 lb./hr. When the system was tested without the precipitator scrubber particle removal efficiencies of 99.5% plus were obtained.

As the installation was designed to process only one-half of the total gas flow from the boiler the results noted above do not represent the actual stack emissions as this consists of both the contribution from the scrubbed flue gas and the particle load remaining in the unscrubbed portion.

An estimate of precipitator efficiency can be obtained from the test results by using the inlet loading to the test series before and after the precipitator as a measure of the inlet and outlet condition of the precipitator. This estimate is shown in Table 17.

The improvement in the efficiency of precipitator operations, as shown in Table 17, as the amount of gas flow to it decreases was confirmed in a later series of tests which showed an increase to 99% efficiency in the ESP when the gas flow was divided equally between the scrubber and the precipitator.

5.2 SO_x EMISSIONS

Results of the sulfur oxides emissions tests are summarized in Table 18 and represent the average of four de-

TABLE 17

ESTIMATE OF ELECTROSTATIC PRECIPITATOR EFFICIENCY

Boiler Avg. Load	Particle Load (Avg.) Inlet	Outlet	Efficiency (Based On 3.6 GR/SCFD Inlet Load	Source (Test Series)
MW	Gr/SCFD	Gr/SCFD		
183	3.654			7
183		0.190	94.8%	5-A
96		0.101	97.2%	5-B

TABLE 18
SO_x EMISSIONS TEST RESULTS
FGD SYSTEM

Test Series	$P^{(1)}$ in H_2O	SO_2 (PPM) Inlet	SO_2 (PPM) Outlet	SO_2 Removal %	SO_3 (PPM) Inlet	SO_3 (PPM) Outlet
5A	15.1	779	78	90	34.6	3.56
5B	6.6	1373	157	88.7		
6	5.1	800	137	82.9	47.5	3.31
7	14.7	1418	88	93.9	2.9	0.64
8	5.2	1419	156	89.0	1.8	0.41

(1) Second stage (absorber) pressure drop.

*Test results abstracted from York Research Corporation, Final Report, Y-8513, Jan. 31, 1975.

terminations per series. Except for tests series 5A most results for SO_2 removal were reasonably close to the average, in test series 5A removal efficiencies measured in the individual tests ranged from 83.5% to 96.8%.

In general, the trends noted by these results (reduced efficiency at lower pressure drop and lower inlet SO_2 concentration) are similar to those seen in the oil fired application of this prototype FGD method. The correlation developed from the New England SO_2 Control Project has been used to analyze the data from this project.

SO_3 inlet and outlet concentrations were also determined during the performance test series and show a significant reduction across the two stage scrubber for both gas taken before the electrostatic precipitator and the gas taken after the electrostatic precipitator cases.

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(1) Maryland State Department of Health and Mental Hygiene 10.03.39 "Regulations Governing the Control of Air Pollution in Area IV", maximum allowable emission of particulate for all solid fuel burning installations greater than 200 million BTU/hr. furnace are 0.03 GR/SCFD.

TABLE 20

COAL ANALYSES - COMPOSITE TO BUNKER #3

DATE	1974						
From	11-1	11-6	11-13	11-21	11-27	12-1	12-11
To	11-5	11-12	11-19	11-26	11-30	12-10	12-17
Total Moisture	6.03	5.93	8.67	7.50	7.95	9.66	10.69
Dry Volatile	22.58	22.31	21.58	21.14	22.44	23.36	25.18
Dry Ash	21.85	20.43	19.11	17.77	21.06	19.23	19.50
Dry S	2.36	2.36	2.15	2.15	2.25	2.13	2.17
Dry Btu	11,684	11,826	12,096	12,399	11,686	11,934	11,911
M&AF Btu	14,951	14,862	14.954	15,078	14,804	14,775	14,796

DATE			1975					
From	12-18	12-25	1-1	1-8	1-15	1-22	2-1	2-5
To	12-24	12-31	1-7	1-14	-121	1-31	2-4	2-11
Total Moisture	12.40	7.90	8.06	8.37	3.18	8.19	8.70	9.78
Dry Volatile	22.99	24.66	24.26	24.74	22.21	25.12	23.67	24.62
Dry Ash	17.53	15.51	17.05	15.34	13.09	17.80	16.68	16.09
Dry S	2.42	2.41	2.42	2.20	2.26	2.06	2.28	1.88
Dry Btu	12241	12711	12467	12809	13492	12477	12553	12514
M&AF Btu	14843	15044	15030	15130	15524	15142	15061	14914

Random Sample		
	Arsenic	13.6 ppm
	Tellurium	1.4 ppm
	Selenium	1.1 ppm

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TABLE 19

REMOVAL EFFICIENCY FOR PARTICLE SIZE RANGES

Test Series	Removal Efficiency		
	Above 5μM	1 to 5 μM	Below 1μM
5 - 1	99.0	97.5	94.6
5 - 2	99.8	99.2	99.0
5 - 3	99.0	99.9	98.6
5 - 4	98.8	97.5	97.0
Average	99.2	98.4	97.3

6.0 PROCESS CHEMISTRY

6.1 FIRST STAGE (PARTICLE CONTROL) OPERATION

While most attention was focused on the SO_2 removal section of the system some work was done to determine the composition of the various important streams in the first stage liquor loop. Tests were conducted during the performance test work carred out in August and September 1974.

These tests involved analysis for a number of chemical species in the streams going to the first pond, the first stage bleed, thickener underflow and overflow, and the sump pump discharge. Components analyzed were calcium, potassium, aluminum oxide, magnesium, chromium, cadmium, titanium oxide, silica, sulfates, sulfites, and chlorine. In addition, the pH and total dissolved solids were also measured.

In order to complete the data and provide some standard of comparision a series of baseline analyses was done for both the inlet and outlet of the pond system and the various system water supplies during a period of limited activity of the FGD system, prior to the inauguration of the performance test work.

The data for both the baseline period and the performance test period have been abstracted for a number of streams and are presented in Table 21. In these tests the baseline data are given where available, while the compositions during the testing period (samples were taken intermittently for the period August 29 through September 21, 1974) are presented as an average for all tests days for the first stage bleed, the pond (Basin #1) inlet, and pond return (Make Up Water to Dilution Tank).

TABLE 21

Chemical Composition at Various Streams in the First Stage - Fly Ash Removal System

Sampling Point		Potassium ppm	Calcium ppm	Magnesium ppm	Iron ppm	Cadmium ppm	Copper ppm	Chromium ppm	Mercury ppb
First Stage Bleed	Avg. All Tests	24.6	7.8	562.0	87.4		.94	1.2	12.7
Inlet to Basin #1	Baseline	n.a.	11.4	78.1	38.0	.01	.23	0.6	2.0
	Avg. All Tests	12.9	8.8	755.7	41.3	.04		0.44	
Make Up Water to Dilution Tank	Baseline	16.4	10.1	83.7	69.2	.01	2.42	0.64	2.0
	Avg. All Tests	10.9	8.8	123.2	.4	.04	.08	.13	4.7
Service Water Supply	Baseline	n.a.	13.3	83.6		.07	.49	0.74	2.0

Table 21 (Cont'd)

Chemical Composition at Various Streams in the First Stage - Fly Ash Removal System

Sampling Point		Sodium ppm	Flouride ppm	SO_4 ppm	Phosphate ppm	Cl ppm
First Stage Bleed	Avg. All Tests	17.0	1.4	4921.0	32.4	537.0
Inlet To Basin #1	Baseline	16.2	1.0	1.0	2.1	224.0
	Avg. All Tests		1.5	2024.4	18.0	
Make Up Water To Dilution Tank	Baseline	21.8	1.0	408.1	16.3	64.2
	Avg. All Tests	10.0	1.9	597.8	7.8	
Service Water Supply	Baseline	45.1	1.0	39.2	2.2	

6.2 PROCESS CHEMISTRY

In an aqueous environment the following series of reactions can be envisioned for SO_2 absorption by pulverized solid MgO:

1. Slaking of MgO

$MgO_{(s)} + H_2O_{(l)} \longrightarrow Mg(OH)_{2(s)}$ (6-1)

$Mg\ (OH)_{2(s)} \longrightarrow Mg(OH)_2\ (Soln)$ (6-2)

$Mg(OH)_2\ (Soln) \longrightarrow Mg^{+2} + 2OH^-$ (6-3)

2. SO_2 Absorption

$SO_{2\ (g)} \longrightarrow SO_2\ (Soln)$ (6-4)

$SO_2\ (Soln) + H_2O \longrightarrow H_2SO_3$ (6-5)

$H_2SO_3 \longrightarrow H^+ + HSO_3^-$ (6-6)

$HSO_3^- \longrightarrow H^+ + SO_3^{-2}$ (6-7)

3. Formation of $MgSO_3$ and $Mg(HSO_3)_2$ (6-8)

$Mg(OH)_2\ (Soln) + H_2SO_3 + (x-2)\ H_2O \longrightarrow$

$MgSO_3.XH_2O$ (6-8)

$x = 3$ or 6

$Mg(OH)_2(Soln) + 2H_2SO_3 \longrightarrow Mg(HSO_3)_2 + 2\ H_2O$ (6-9)

$MgSO_3\ x\ H_2O \longrightarrow Mg^{+2} + SO_3^{-2} + x\ H_2O$ (6-10)

$Mg(HSO_3)_2 \longrightarrow Mg^{+2} + 2HSO_3^-$ (6-11)

4. Other reactions (side):

$CO_2\ (g) + H_2O \longrightarrow H_2CO_3\ (Soln)$ (6-12)

$H_2CO_3\ (Soln) \longrightarrow H^+ + HCO_3^-$ (6-13)

$HCO_3^- \longrightarrow H^+ + CO_3^{-2}$ (6-14)

$2\ MgSO_3\ x\ H_2O + O_2 \longrightarrow 2\ MgSO_4 + x\ H_2O$ (6-15)

$Mg(OH)_2(Soln) + H_2CO_3 + 3H_2 \longrightarrow MgCO_2.5\ H_2O$ (6-16)

$Mg(OH)_2(Soln) + SO_3 \longrightarrow MgSO_4 + H_2O$ (6-17)

6.2.1 Absorber Reactions

In the system the main reaction in the second or SO_2 removal stage is the neutralization of H_2SO_3 by hydrated MgO to form insoluble hydrated $MgSO_3$. $MgSO_3$ crystallizes out with either three or six molecules of water. The nature of the hydrate, i.e., hexa- or tri- is partially dependent upon the temperature in the absorber. (Hexahydrate is stable at room temperature while the trihydrate is stable above 108°F). The transition temperature of the two hydrates, according to the literature, is 42.5°C (108.5°F). During operations at PEPCO the temperature of the recycle slurry averaged about 110°F. Since this temperature in the absorber is higher than the transition temperature it was expected that the trihydrate of magnesium sulfite would be the main product as was the case during the work on the Boston Edison Project. However, in actual operation, at Dickerson Station the centrifuge cake analysis showed almost 95% $MgSO_3 . 6H_2O$.

6.2.2 Formation of Oxysulfates

At PEPCO feed MgO was digested in mother liquor that has as its main consitutent magnesium sulfate in solution (approximately 15% dissolved $MgSO_4$).

It is known that MgO may react with $MgSO_4$ solution to form a jell like substance. This jell consists mainly of basic magnesium sulfate, represented as: $xMgSO_4\ yMg(OH)_2\ zH_2O$. The values of x, y, and z are dependent upon the concentration of reactants and temperature.

During early operations at PEPCO, when virgin MgO was used in the unheated mixing tank considerable jell and lump formations occurred. X-ray and chemical analysis of these products showed the presence of $3Mg(OH)_2MgSO_48H_2O$ (i.e., oxysulfate in varying amounts admixed with quantities of unreacted $Mg(OH)_2$.

Laboratory studies, which were initiated to investigate this problem concluded that:

1) The problem appeared to be confined to operation with virgin MgO particularly at high $MgSO_4$ concentrations (15% $MgSO_4$ or more) in the centrate. The problem is further aggravated when the MgO-$MgSO_4$ mixture is allowed to stand for a time with insufficient agitation.
2) When calcined MgO (regnerated) of bulk density above 20 lb/ft^3 is used in the reaction no major problem was observed as long as proper agitation and temperature of digestion of virgin MgO were maintained.

Because of lack of operating time the laboratory observations with regnerated MgO could not be completely confirmed in the plant.

6.3 MAGNESIUM SULFITE CHEMISTRY

Fundamental questions in the Chemico Basic Magnesia Slurry Process are the chemistry of the formation and dehydration of the hexa and tri-hydrate forms of magnesium sulfite, and the mechanism of $MgSO_3$-$MgSO_4$ thermal decomposition in the presence of carbon and a reducing atmosphere.

Understanding of the first would allow control of the production of the hydrate type formed in the absorption system thereby affecting economies in the separation and drying of these materials.

Knowledge of the second might lead to process improvements for production of richer SO_2 off-gas, with more reactive MgO, or direct conversion to elemental sulfur.

Because of the complexity of these questions research work in the three areas was divided between the Chemico laboratory facilities at the Rumford, R.I., plant and the New Jersey Institute of Technology, (N.J.I.T.) Department of Chemical Engineering and Chemistry. Most of the instrument studies, i.e., Mass Spectroscopy, I.R. Spectroscopy, Differential Thermal Analysis, Thermogravimetric Analysis, etc. were performed at N.J.I.T. The results of these studies are given in the Appendicies.

The following section is a summary of the principal findings resulting from the work.

6.3.1 Conditions Governing the Formation of Magnesium Sulfite Hydrates

A literature survey showed the existence of only two hydrates of magnesium sulfite: $MgSO_3.6H_2O$ and $MgSO_3.3H_2O$. The hexahydrate is more stable at ambient conditions. The hydration energy of the two are as follows:

$$MgSO_3 + 6H_2O \longrightarrow MgSO_3.6H_2O \text{ -13.38 K Cal.}$$
$$MgSO_3 + 3H_2O \longrightarrow MgSO_3.3H_2O \text{ -11 k Cal.}$$

The N.J.I.T. study of the reaction of SO_2 (at 320°F) with MgO slurry at 120°, 135°, 149° and 176°F showed that at 120° and 135°F the major product was $MgSO_3.6H_2O$ while the trihydrate, $MgSO_3.3H_2O$ was the dominant product at 149° and 176°F. Presence of $MgSO_4$ in solution (up to 18%) and passage of excess air did not have any significant influence on the nature of the hydrate formed.

At Rumford, the reactions of SO_2 (at 300°F) with MgO (virgin and regenerated) were studied at 135°, 140-145° and 150°F. The pH was maintained at 6.8-7.20. Samples were collected periodically and the filtrates were returned to the reactor. The results of the study agreed with the conclusions noted above.

At 135°F the major product was the hexahydrate and trihydrate was the predominant one at 150°F. The study at 140-145°F was interesting--it showed the gradual change from hexahydrate (original product) to trihydrate (final product). The complete transormation of the hexahydrate to trihydrate took place in 4-5 hours of run. The study (140-145°F) also showed that the transition of hexahydrate to trihydrate is accompanied by a reduction of crystal size (from 200 to 50 micrometers).

Analysis of centrifuge cakes from PEPCO and Boston support the validity of the contention that (a) the ratio of the two hydrates at any temperature is controlled by the equilibrium value at that temperature, (b) hexahydrate persists even above the transition temperature and (c) the rate of transition is slow (cf. Kovachev Ts. et al., Khimiya i Industriya. Sofia 42 (5) (1970) 209-211. Transition temperature of the hydrates, 40°C or 104°F).

Centrifuge Cake From
Boston (135^{o}F)
2-20-74 to 2-21-74

Hours	Ratio of $\frac{3H_2O}{6H_2O}$
0	0/100
4	12.5/87.5
12	30/70
16	37/63
20	49/51
24	55/45
28	58/42
32	70/30

Centrifuge Cake From
PEPCO (120^{o}F)
12-10-73 to 2-15-73

Hours	Ratio of $\frac{3H_2O}{6H_2O}$
24	16/84
48	9/91
74	11/89
97	18/82
130	7.5/92.5

The variation in PEPCO samples was probably due to temperature fluctuations in the system (the PEPCO system did not stablize at the time. However, analysis of centrifuge cakes at later dates always showed a preponderance of hexahydrate in PEPCO cakes).

6.3.2 Dehydration of Magnesium Sulfite Hydrates

DTA, TGA and DSC studies with the two hydrates of magnesium sulfite, $MgSO_3$ $6H_2O$ and $MgSO_3$ $3H_2O$ showed that the path of dehydration and temperature of dehydration are dependent on the manner of heating, i.e., whether dehydration took place in an enclosed vessel or open or flow-through vessel. (Appendix 2).

For example:

	CONDITIONS	
DTA Study	CLOSED	OPEN
$MgSO_3$ $6H_2O$	Two endotherms observed (at 107^{o} and 205^{o}C)	One endotherm only (at 90^{o}C)

CONDITIONS

DTA Study	CLOSED	OPEN
$MgSO_3$ $3H_2O$	One endotherm only (at 205°C)	One endotherm (at 160°C)

TGA study also showed similar behavior of the two hydrates:

	CONDITIONS OF STUDY	
TGA Study	CLOSED	OPEN
$MgSO_3$ $6H_2O$	Two endotherms (at 125° and 220°C)	One endotherm only (at 70°C)
$MgSO_3$ $3H_2O$	One endotherm (at 200°C)	One endotherm (at 100°C)

This study also showed that TGA (enclosed environment) can be used to determine quantitatively the amounts of the two hydrates in a mixture of the two. Presence of inerts (any substance which does not change weight between 175° and 400°C) does not interfere with the determination. The only potential interferring element relevant to magnesia process is $MgSO_4$ $7H_2O$ ($-7H_2O$ at 200°C). The interference by $MgSO_4$ $7H_2O$ can be easily avoided by washing the sample with absolute ethyl alcohol (only $MgSO_4$ &H_2O dissolves). The amounts of the two hydrates, in a mixture, are calculated as follows:

$$\%MgSO_3\ 6H_2O = \frac{\%\ \text{wt. loss at } 175^{\circ}\text{C} \times 2 \times 100}{50.9}$$

$$\%MgSO_3\ 3H_2O = \frac{100 \times (\%\ \text{wt. loss at } 400^{\circ}\text{C} - \text{wt. }\%\text{ loss at} 175^{\circ}\text{C})}{34.10}$$

Additional dehydration studies were carried out in the laboratory.

The two hydrates were heated in a vacuum oven (12 inches of mercury) at $80^{\circ}C$ ($176^{\circ}F$), $120^{\circ}C$ ($248^{\circ}F$) and $150^{\circ}C$ ($302^{\circ}F$) for extended periods of time. Samples were withdrawn at intervals and chemically analyzed.

At $176^{\circ}F$ and $248^{\circ}F$ little or no change was observed with $MgSO_3\ 3H_2O$ even after 25 days. At $176^{\circ}F$ the hexahydrate changed to trihydrate after 2 days (confirmed by chemical and x-ray analysis). At $248^{\circ}F$ the hexahydrate dehydrated to yield a hydrate of apparent composition $MgSO_3 2H_2O$ after 21 days.

At $302^{\circ}F$ the hexahydrate changes to trihydrate within two hours. Holding the compound at this temperature results in a continuous, gradual loss of water. After 200 hours at $302^{\circ}F$ the starting hexahydrate had an apparent composition of $MgSO_3 \frac{1}{2}H_2O$. When starting with the trihydrate it appears to be stable up to 48 hours at the $302^{\circ}F$ temperature. Then it looses water rapidly to give the apparent hydrate $MgSO_3 \frac{1}{2}H_2O$.

The rates of dehydration of the two hydrates at $330^{\circ}F$ are also different the hexahydrate loosing water more rapidly than the trihydrate. The rates of dehydration at $330^{\circ}F$ were studies using Ohaus moisture balance. To form the anhydrous salt, $MgSO_3$, containing 10% moisture at $330^{\circ}F$ following retention periods were necessary:

for	$MgSO_3 6H_2O$	10 minutes
	$MgSO_3 3H_2O$	25 minutes

The dehydration study indicates that the water molecules are more strongly bound in the trihydrate than in hexahydrate.

6.3.3 Mass Spectroscopy of the Two Hydrates of Magnesium Sulfite vs. Temperature

The two hydrates were heated from 35° to $400^{\circ}C$ in a

Finnigan Quadrupole 1015 mass spectrometer. No other species except SO_2 and H_2O were observed. The plot of log $\left\{\frac{SO_2}{H_2O}\right\}$ vs. temperature showed that the two hydrates decomposed in a similar manner (under vacuum conditions). (Appendix 3).

6.4 PULVERIZATION

The SO_2 removal capability of MgO in the Chemico magnesia system is dependent on both intensive and extensive properties of the regenerated material.

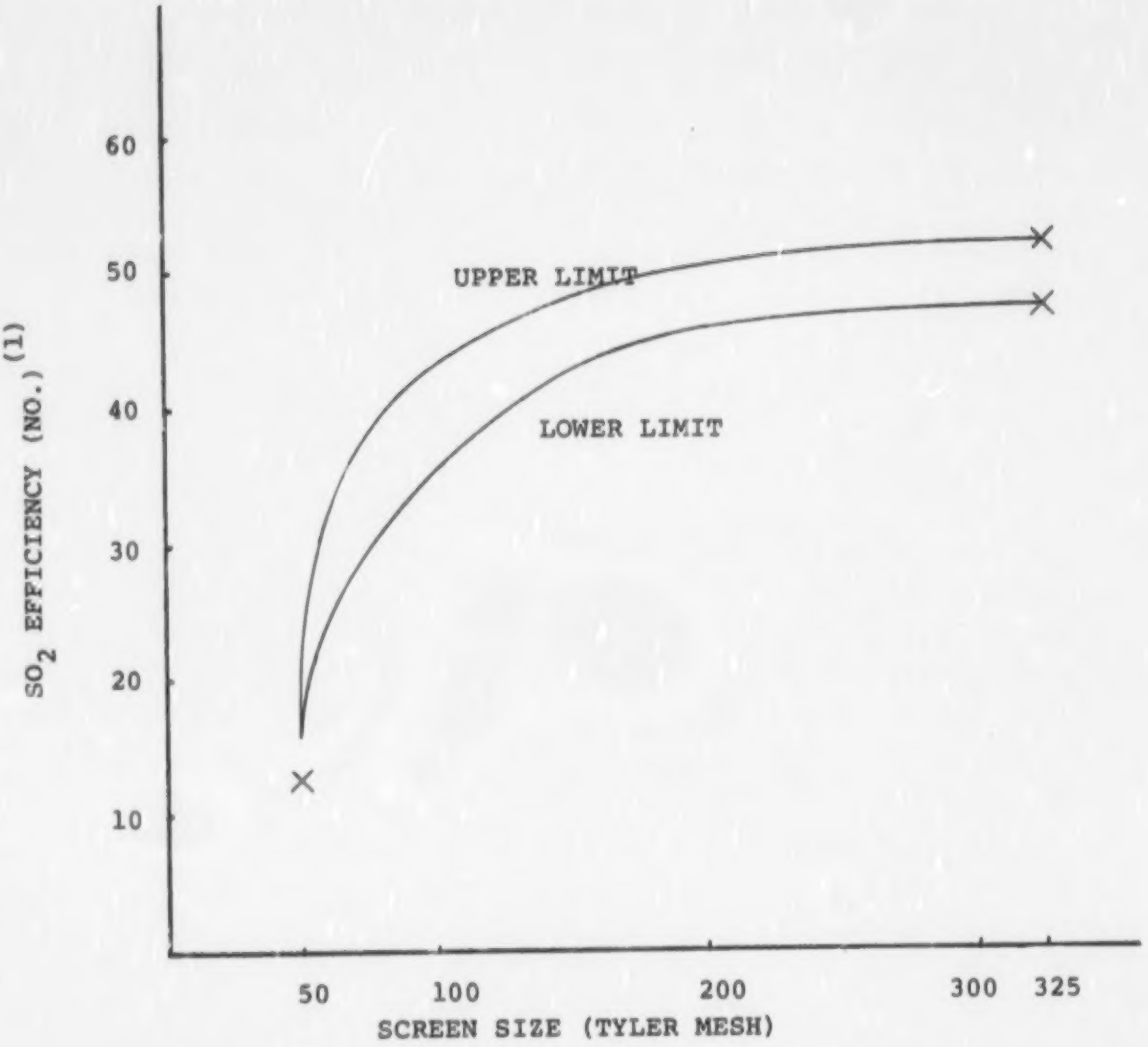
Of the extensive properties, the size range of the MgO particles returned for slaking is very important. Experimental data indicate that grinding the calcined magnesia increases its reactivity. However, there is a limit to which that reactivity can be increased.

Laboratory data now indicate that a size range of -100 to +200 mesh (between 75 and 150 micrometers) is satisfactory for recycled, regenerated magnesia. This is shown in the attached graph (Figure 8). The greatest increase in MgO activity occurs when +50 mesh size particles are reduced to -50 to +100 mesh, additional improvement in SO_2 removal is obtained when size is reduced to -100 to +200 mesh. Reduction to finer size has only a slight affect on activity increase.

The properties of regenerated MgO (shown in Table 22) are determined by both calcining conditions and contaminent level.

FIGURE 8

EFFECT OF PULVERIZATION ON SO_2 EFFICIENCY



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(1) Measure of uptake of SO_2 by measured quantity of regenerated MgO by the method described in "The Magnesia Scrubbing Process as Applied to an Oil Fired Power Plant" EPA 600/2-75-057 pg. 253 ff.

TABLE 22

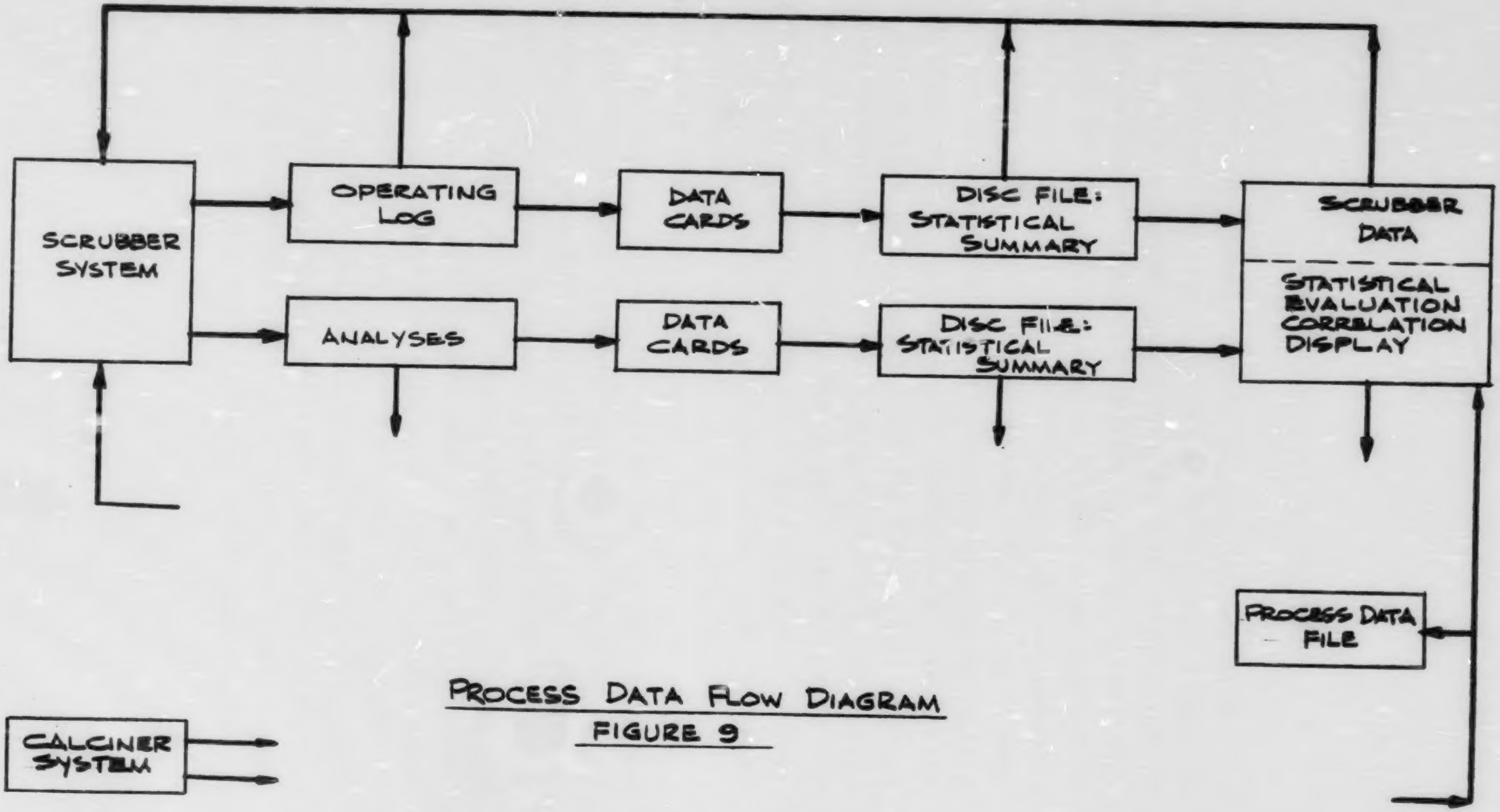
PROPERTIES OF DIFFERENT FRACTION REGENERATED MgO
(SAMPLE NO. PR-289)

Analysis	+50 Mesh Fraction	-50 to +100 Mesh Fraction	-100 to +200 Mesh Fraction	-200 to -325 Mesh Fraction	Ground +50* Mesh Fraction to (-100 to +200) Mesh
%MgO	89.96	89.46	90.32	91.73	
%$MgSO_4$	1.73	2.43	2.60	2.35	
%$MgSO_3$	0.50	1.42	1.33	0.98	
SO_2 Eff.	13	37	47	48	48

7.0 CORRELATION OF PROCESS DATA

With the initiation of the planned operations program process data were collected and recorded regularly. In order to assimilate, store, analyze and deseminate this information efficiently, a computerized method of handling the data was employed. As a result, it was possible to effect a real-time feedback to the process or correlated and trend results at the time they are most useful, along with accumulating the information in a permanent data bank available for analysis and retrieval. The statistical computer program package utilized, permits the use of the same format data file as input to a wide range of sub-programs, including regression calculations, x-y plotting, file listings, trend plots, etc. This flexibility reduced the number of files that needed to be maintained, and provided speed advantages.

Figure 9 illustrates the flow of information between scrubber and calciner operating systems and the computer data bank. Operating conditions and analyses are entered on punch cards on a daily basis for primary storage and transferred to computer disk files for futher processing. At monthly intervals the following statistics were generated for all variables: average, maximum, minimum, standard deviation, and a percent change relative to a base period. The final stage in the data flow was the integration of the operating log and analysis files, plus the inclusion of appropriate time lags for the scrubber and regeneration facilities to form single disk files suitable for input to correlation and plotter computer programs. The process data bank becomes the source of retrieval of process information for continued analysis and for comparison with


SCRUBBER SYSTEM
OPERATING LOG
ANALYSES
DATA CARDS
DATA CARDS
DISC FILE: STATISTICAL SUMMARY
DISC FILE: STATISTICAL SUMMARY
SCRUBBER DATA
STATISTICAL EVALUATION CORRELATION DISPLAY
PROCESS DATA FILE
CALCINER SYSTEM

PROCESS DATA FLOW DIAGRAM

FIGURE 9

results from other installations.

The process correlations developed in the previous "New England SO_2 Control Project", EPA Ref. CPA 70-114, were developed primarily using the data bank as input to a stepwise multiple linear regression program. This approach minimizes the number of computer runs required to determine equations that are statistically significant, prior to their evaluation for consistency with observed data. F ratio tests are employed to establish significance levels for testing of equation variables, and coefficients are calculated by least squares techniques. With respect to the SO_2 removal efficiency correlation, log transformations were required by the regression program. Continued investigation of mass transfer coefficient in this program may lead to the use of non-linear programs.

Normal excursions in the process variables are sufficient to satisfy the required scope of the test program variation in levels of operating data to fulfill the requirements of this method of operations data analysis.

7.1 SO_2 REMOVAL

A correlation relating SO_2 removal efficiency, pressure drop and SO_2 inlet concentration had been developed during the previous operation of the MgO system prototype on the oil fired boiler. Included in the formulation of the correlation were data points from the performance testing of Dickerson. The data accumulated during the additional operation of this project was used to improve the correlation based on the variations of SO_2 removal efficiency and other operating factors which were observed.

7.1.1 First Stage SO_2 Absorption

A series of tests were performed with the scrubber to assess the SO_2 removal capability of the first (particle removal) stage of the system. Results of these tests are given in Table 23. These tests showed an average removal efficiency for the first stage of 6.7%.

7.1.2 Effect of Pulverization of Regenerated MgO

During the period of scrubber operation at PEPCO from November 1974 to January 1975, the observed SO_2 removal efficiency varied more than could be expected from normal changes in operating conditions, i.e., pressure drop, SO_2 inlet concentration, etc. Efficiency decreased approximately 10% in November before recovering in December and January. The efficiencies were adjusted to constant operating conditions and plotted to show that a significant variation in SO_2 removal remained after adjustment of pH and pressure drop which was assumed to be related to other process variables. Figure 10 is plot of removal efficiency vs. time illustrating this phenomenon. The scale of the plot is expanded to show the effect more clearly.

Laboratory studies indicated that a critical particle size range was required to maintain activity of the regenerated magnesia. In order to determine whether any relationship existed between MgO particle size distribution and SO_2 removal activity in the FGD system, a series of MgO samples were taken from the weigh feeder at Dickerson for screen sizing and activity measurements. Results of this analysis are shown in Table 24 along with corresponding scrubber operating conditions for the time frame over which the samples were taken.

This information was analyzed by linear regression techniques and showed a significant particle size correlation with SO_2 removal efficiency in the -50 to +100 mesh range. The correlation is given in equation 7-1.

$$R = -7.68 - 0.0338 (P_{100})^2 \qquad (7\text{-}1)$$

R = % reduction in SO_2 removal efficiency

Where P_{100} = % (-50 to +100 mesh) material in MgO Feed

TABLE 23

SO_2 REMOVAL IN FIRST STAGE

Date	Time	SO_2[(1)] Inlet ppm	SO_2[(1)] Outlet ppm	Efficiency %
1/8/74	1700	1300	1220	6.2%
	1900	1300	1290	0.8
	2100	1390	1200	13.7
	2300	1430	1340	6.3
1/9/74	2100	1300	1300	0.0
	0300	1390	1360	2.2
	0500	1500	1380	8.0
	0700	1300	1300	0.0
	0900	1070	990	7.5
	1100	1080	900	16.7
	1300	980	900	8.2
	1500	900	800	11.1

Average Removal Efficiency 6.7%

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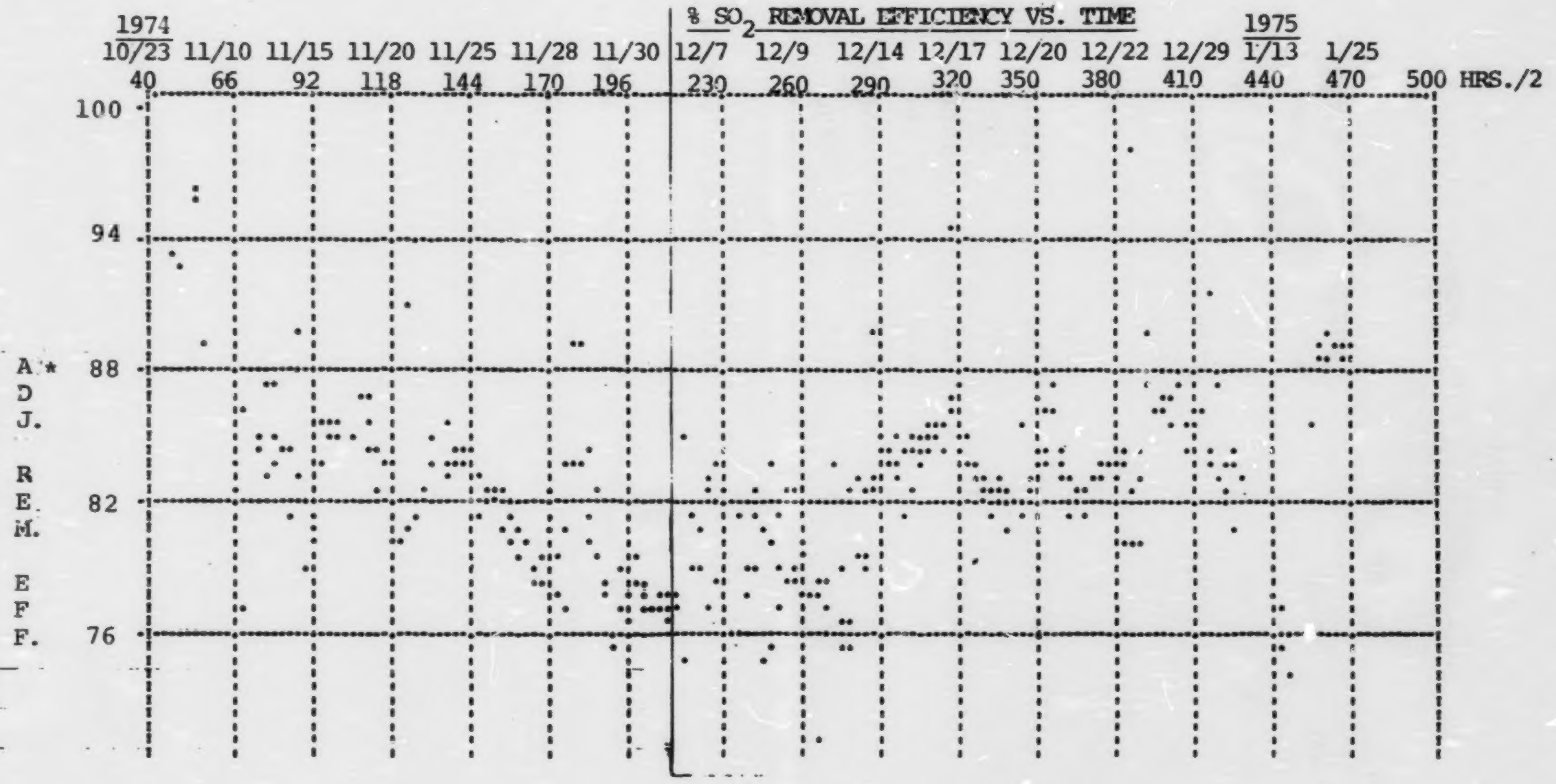
(1) Abstracted from the continuous emissions monitoring data obtained using a Flourescent SO_2 analyzer.

FIGURE 10



*Observed SO_2 rem. eff. adjusted to common basis of 9"Δ P, 7 pH 1900 PPM Inlet SO_2.

TABLE 24

WET SCREEN ANALYSIS OF MgO BELT SAMPLES

Date (1974)	11/20	11/21	11/24	11/25	12/1	12/9	12/18	12/20	12/23	12/27
Time	---	---	---	---	---	0800	2000	1200	0400	1200
Tyler Screen Mesh Size - %										
+50	1.0	0.8	0.5	0.3	4.6	5.2	3.6	6.6	0.7	1.8
-50,+100	0.8	0.5	0.6	0.8	14.5	5.4	5.7	6.8	1.9	2.5
-100,+200	2.6	1.5	1.3	1.7	8.2	12.2	10.3	13.3	7.7	3.6
-200,+325	4.1	2.2	2.7	3.3	6.9	11.8	13.8	10.1	13.6	2.5
-325	91.5	95.0	94.9	93.9	65.8	65.4	66.6	63.2	76.1	89.6
Operating Conditions										
P, In. H_2O	(1) 9.5	(1) 9.3	(1) 8.9	(1) 8.2	(1) 6.4	4.2	8.4	3.6	8.5	5.9
SO_2 Removal Eff., %	85.1	86.2	83.8	84.1	76.0	79.0	82.0	80.6	83.6	83.0
SO_2 Inlet Conc., PPM	1210.	1240.	910.	920.	1080.	1170.	1200.	960.	840.	900.
Predicted (2) SO_2 Removal Eff. - %	85.0	85.0	84.9	84.6	76.4	79.5	83.5	77.0	84.6	82.9
(Actual-Predicted) Efficiency, %	0.1	1.2	-1.1	-0.5	-0.4	-0.5	-1.5	3.6	-1.0	0.1

(1) Average Conditions For Day

(2) From Equation 7-1

After determining the pulverized MgO particle size the resultant correction can be subtracted from the efficiency obtained from the correlation developed from the previous operations. This basic correlation is given in equation 7-2.

$$\text{SO}_2 \text{ Removal Efficiency (Base)} = 1-\text{Exp}\left(\left(2.666\,(\Delta P)^{A}\,(\text{SO}_{2I})^{B}(10)^{C}-3\right)\right) \qquad 7\text{-}2$$

Where $A = -1.014$

$B = -3.75 + 0.271 \ln \text{SO}_{2I}$

$C = 6 - 0.031\ \text{pH}$

SO_{2I} = SO_2 inlet concentration (ppm)

ΔP = pressure drop, inches H_2O

Figure 11 is a plot of the predicted effect of the concentration of -50, +100 mesh size in the MgO on SO_2 removal using equation 7-1 at constant process conditions. Results show that SO_2 removal decreases with increasing concentration of this size range, which is consistent with laboratory observations. Figure 12 is a plot of observed vs. predicted efficiency during the subject time period.

7.1.3 Reduction in Activity of Magnesia

Equation 7-1, showing particle size effect also predicts a 7.68% offset from removal efficiency calculated from 7-2, a correlation based on ΔP, pH, and inlet SO_2 concentration.

An analysis of inventory records during the period covered by the testing program revealed that the magnesia feed consisted of a mix of virgin and regenerated MgO. The beginning inventory on October 18th at Dickerson was 46 tons of regenerated MgO in the following week 140 tons of virgin MgO were added as makeup to the silo. About 22 tons of regenerated MgO were returned to Dickerson in the period from October 18th through November 29th while about 120 tons of MgO were consumed in the operations there. Thus during the period of reduced efficiency virgin material was the principal feed to the system. A comparison of the manufacturer's analysis, Table 25,

FIGURE 11

EFFECT OF MgO PARTICLE SIZE ON SO_2 REMOVAL EFFICIENCY

BASIS: PEPCO Operation Nov.-Dec. 1974

Adjusted to 7" Δ P, 7 pH, 1000 PPM SO_2 Inlet

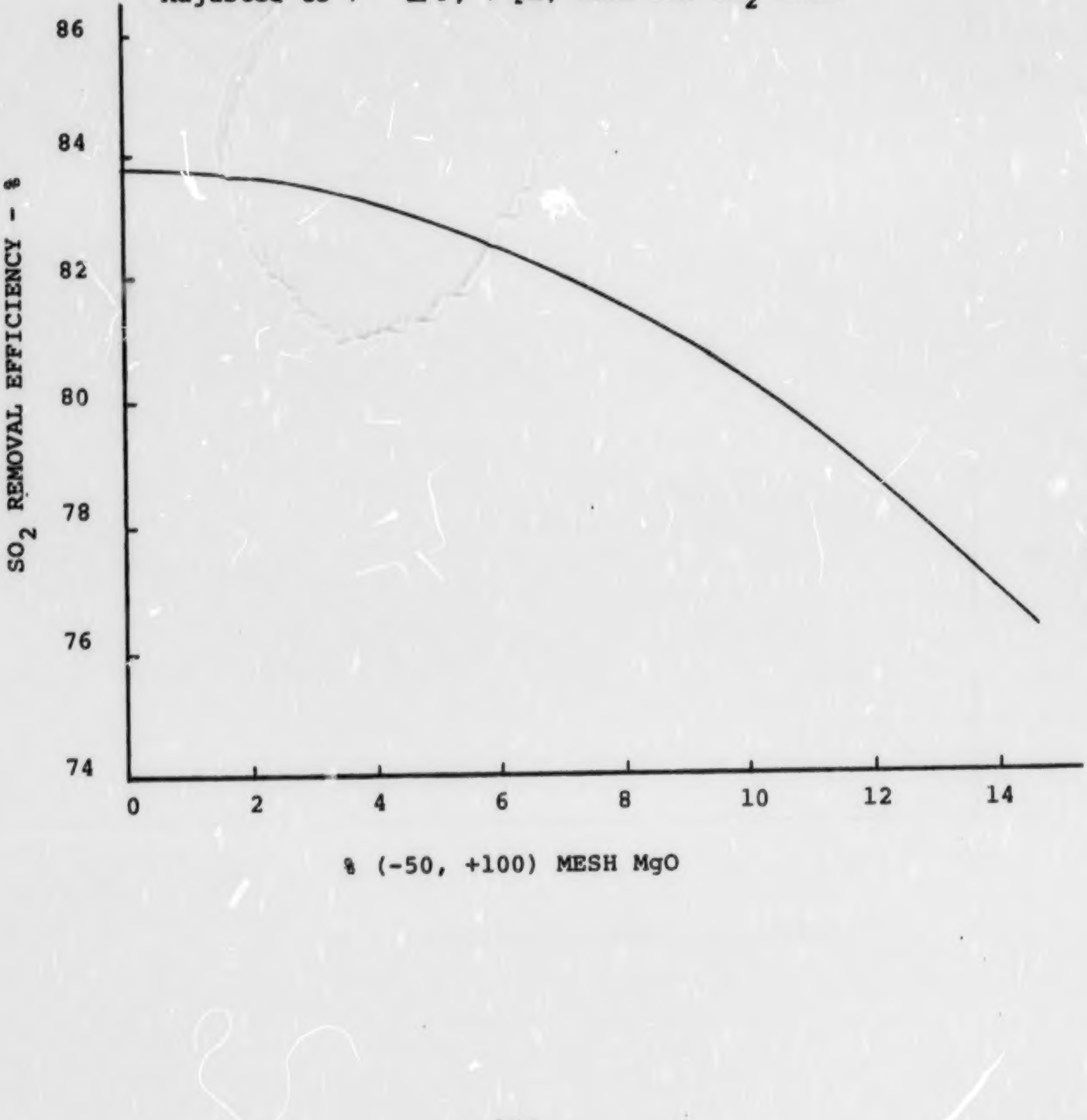


FIGURE 12

OBSERVED VS. PREDICTED SO_2 REMOVAL EFFICIENCY

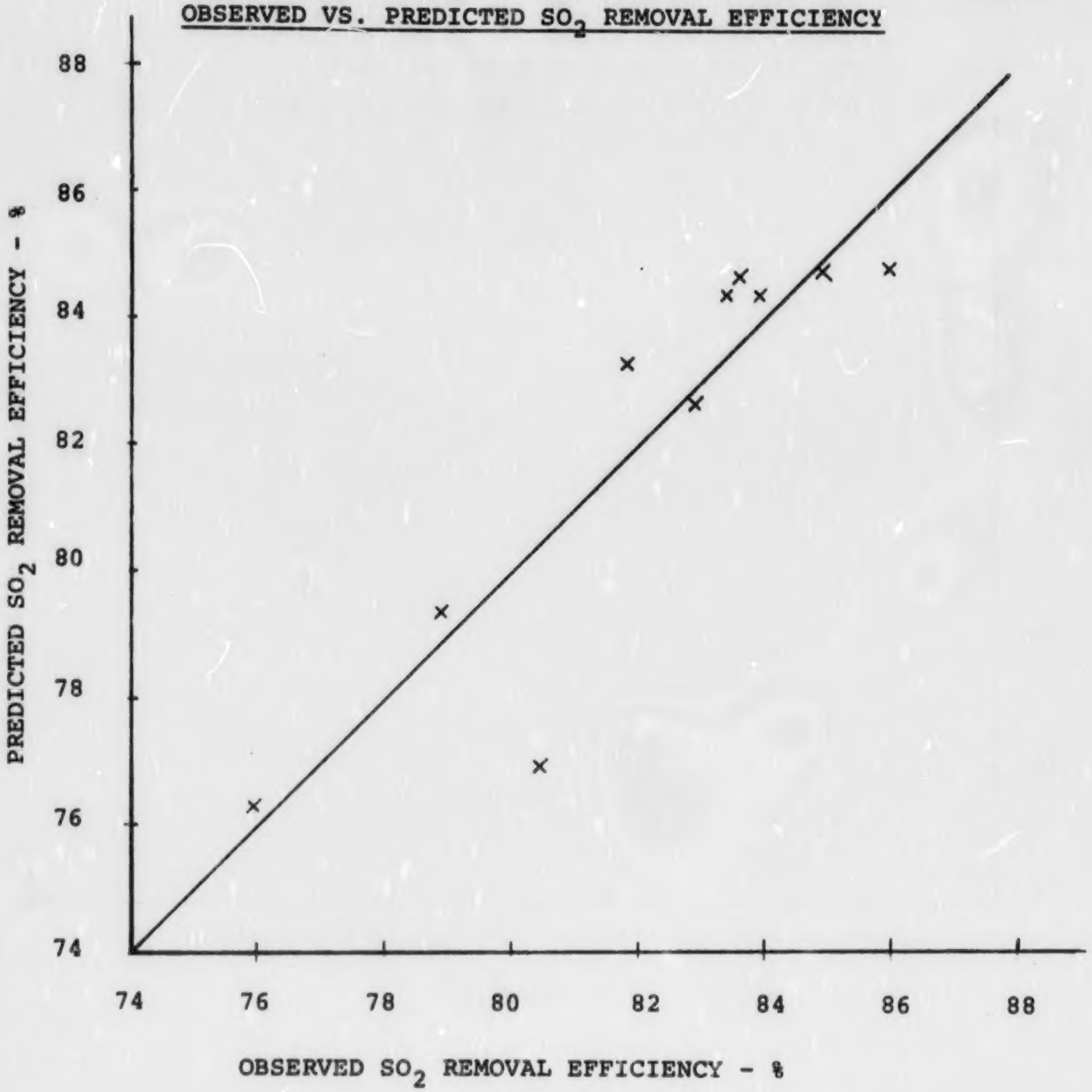


TABLE 25

MANUFACTURERS ANALYSIS OF MAGNESIA SHIPPED TO DICKERSON STATION

	7/12/74 L&N200196	8/20/74 L&N200487	10/7/74 L&N200257	10/25/74 L&N200304
CaO	1.62	1.78	1.94	1.96
SiO_2	0.7	0.60	0.61	0.62
R_2O_3	0.51	0.54	0.43	0.70
Cl	0.18	0.26	0.12	0.19
SO_3	0.69	0.78	0.67	0.64
-325 M	99.75	99.09	98.71	99.61
LOI	6.22	5.46	4.33	4.57
S.A.	66.7	52.4	79.3	94.7

Where R_2O_3 = Iron and aluminum oxides
-325 M = Percent less than 325 mesh
LOI = Loss on ignition
S.A. = Surface area

for the virgin MgO supplied during the period of the program showed little deviation from that supplied at the beginning of the Development Program.

It has been noted (Sec. 6.3.4) that analysis of materials from the MgO slurrying system in the earliest phase of the project showed that the highly reactive virgin MgO exhibited a stronger tendency to form magnesium oxysulfates, which would hinder the SO_2 absorption efficiency of the magnesia slurry. It is thought that the deviation noted here is a measure of the additional reduction in efficiency resulting from the formation of the complex when a feed of virgin MgO is admixed with centrate.

7.2 CENTRIFUGE OPERATION

Centrifuge operations at both PEPCO and Boston Edison were compared and analyzed with regard to process optimization. The effect of centrifuge feed rate on efficiency and solids removal rate was determined by regression analysis of the operating data. Since the two centrifuges are mechanically identical, it would be expected that similar operating characteristics would be observed, and this was confirmed by analysis. The following results were obtained:

1. Centrifuge efficiency, i.e., pounds solids removed in centrifuge cake per pound solids in centrifuge feed, was shown to be a function of centrifuge feed rate with efficiency declining with increasing throughput.
2. Solids removed from the system was shown to pass through a maximum value as total feed rate to the centrifuge is increased.

Figures 13 and 14 show these results for centrifuge operation with average recycle compositions, normal rotational speed and standard weir height.

FIGURE 13

EFFECT OF CENTRIFUGE FEED RATE ON SEPARATIONAL EFFICIENCY

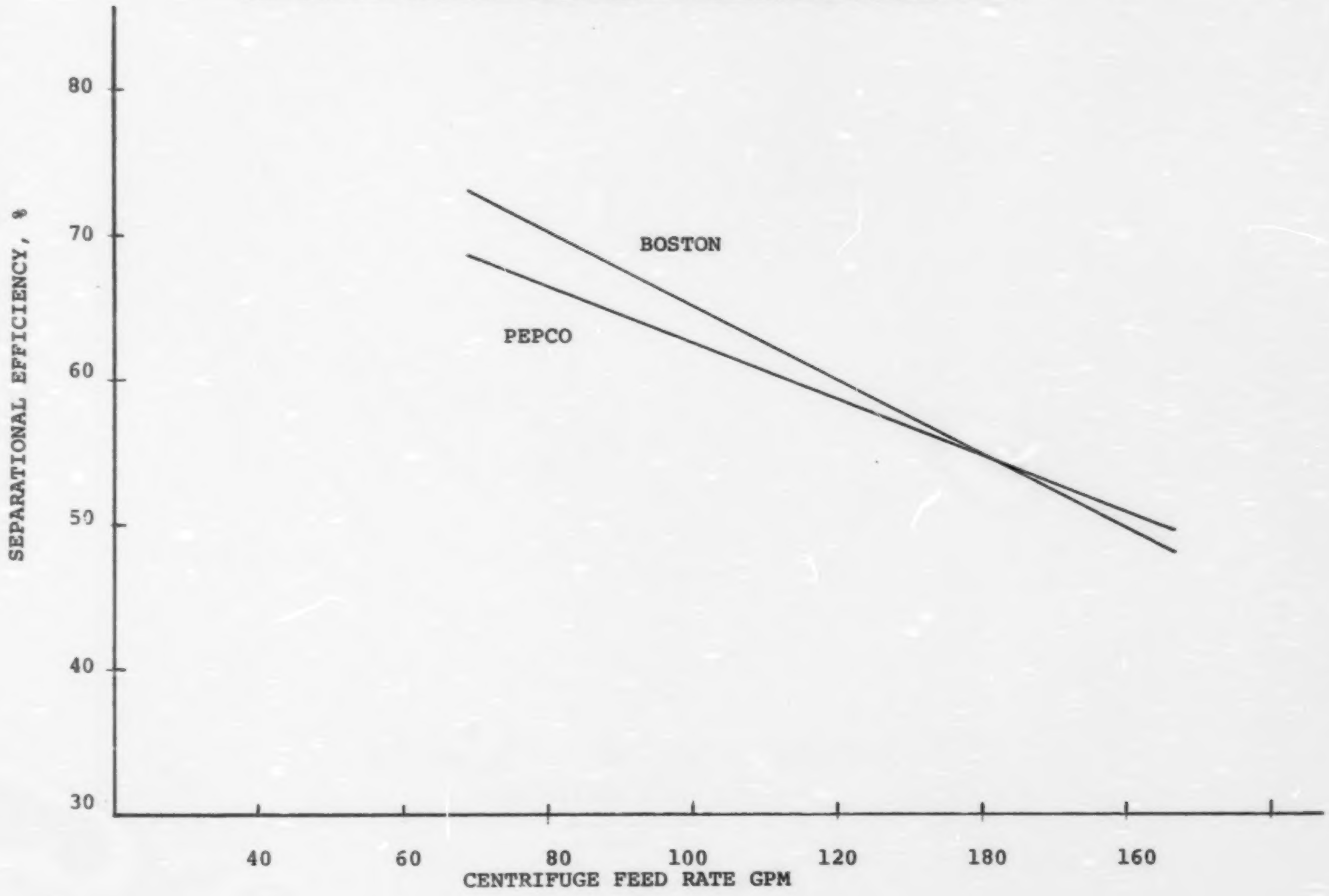
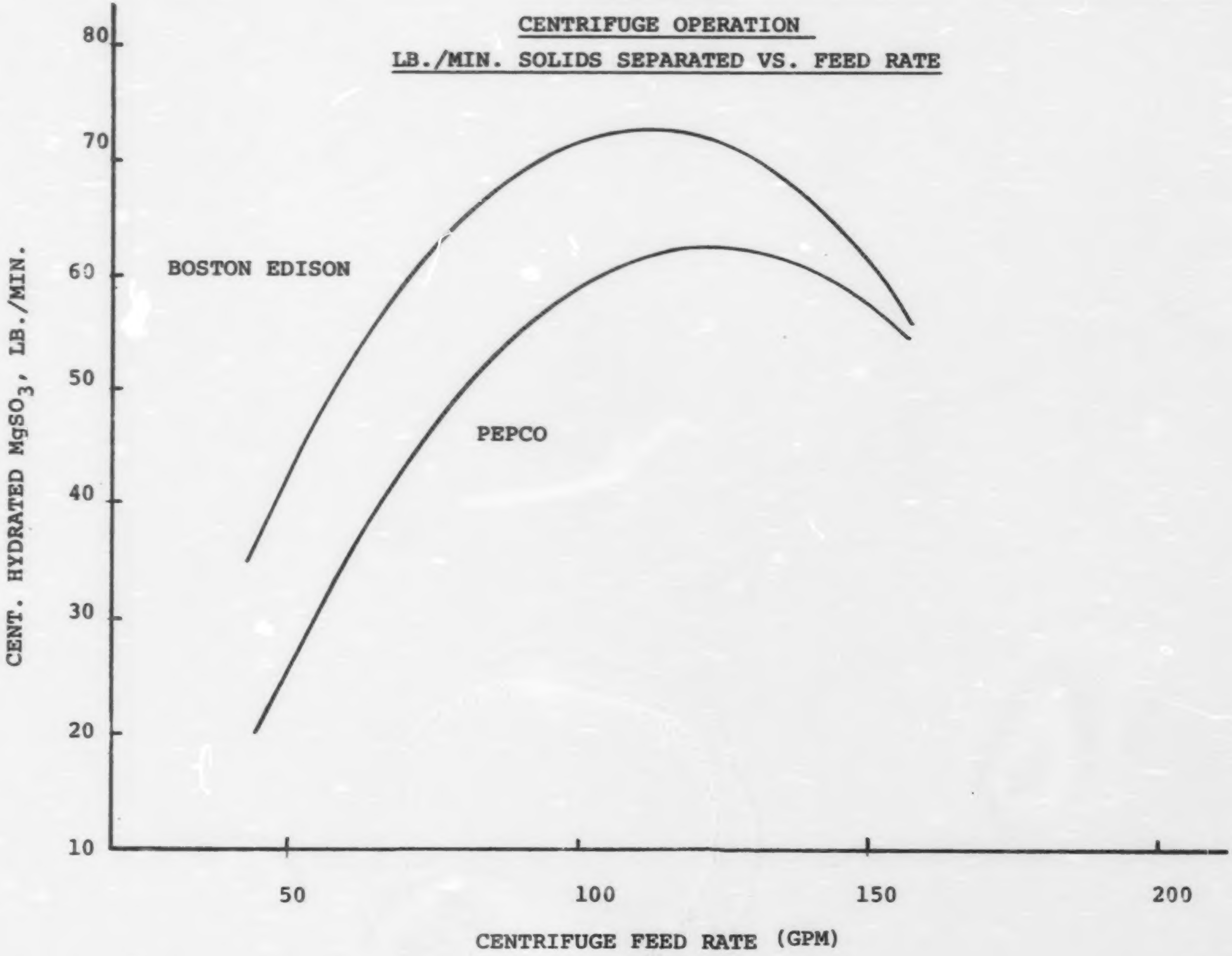


FIGURE 14

CENTRIFUGE OPERATION

LB./MIN. SOLIDS SEPARATED VS. FEED RATE



7.3 DRYER OPERATION

Data from the operations at PEPCO were analyzed to determine the influence of dryer operation conditions on dryer efficiency. A scatter plot of dryer solids vs. product temperature is shown in Fig. 15. In addition, the data collected during the Boston Edison operation were analyzed and compared. The results of preliminary review of the effect of dryer product and dryer outlet gas temperature on product moisture content is shown in Fig. 16. Variations in the performance of the dryers is attributed to the use of a counter-current unit at Boston and a co-current unit at PEPCO. A statistical analysis of this data shows a minor interaction between dryer product and dryer outlet gas temperature.

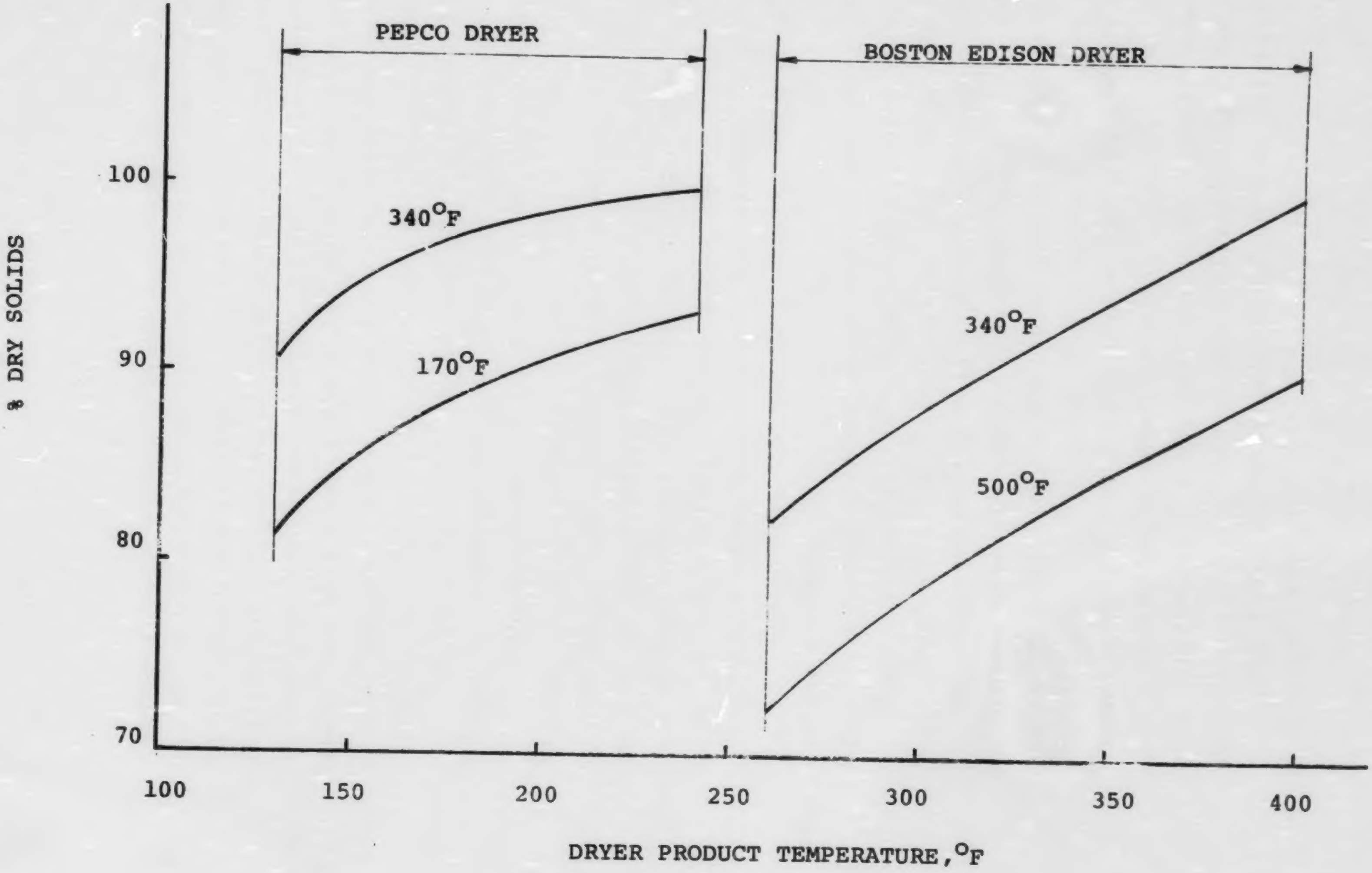
An analysis of the change in the ratio of $MgSO_4$ to $MgSO_3$ in the dryer feed and product streams, an indication of oxidation accurring in the dryer, showed a slight decrease in the $MgSO_4$ concentration for PEPCO operations as compared to the oil fired boiler application for the average range of system operations. This is shown in Figure 17.

7.4 CALCINER FEED RATE

A limitation on reliable operation of the integrated plants was reduced calciner feed rate. A low (40 lb/min.) feed rate was necessary to control slides and unprocessed $MgSO_3$ in the calcined product, Fig. 18. Initial investigations during the work at Mystic Station indicated a relation between particle size of the dryer product and frequency of slides. Further investigation showed that variation in size distribution of the dryer product was associated with variation in moisture content of the product. During the final phases of the program at Boston the dryer was operated to produce a product with a moisture content of approximately

FIGURE 15

COMPARISON OF PEPCO AND BOSTON EDISON DRYER OPERATION
(FOR RANGES OF DRYER GAS OUTLET TEMPERATURE)



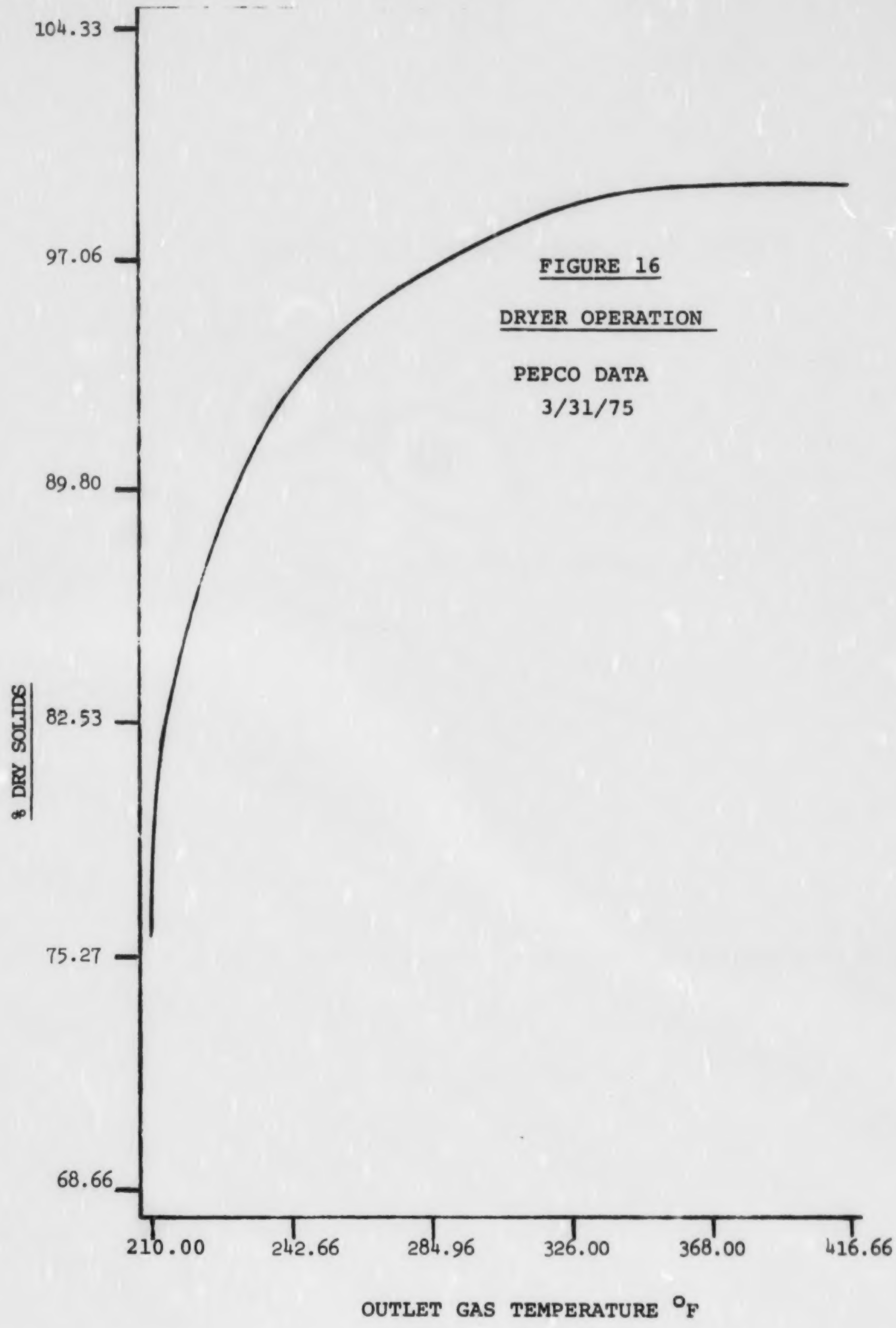


FIGURE 16

DRYER OPERATION

PEPCO DATA
3/31/75

FIGURE 17

COMPARISON OF $MgSO_3$ OXIDATION IN DRYER

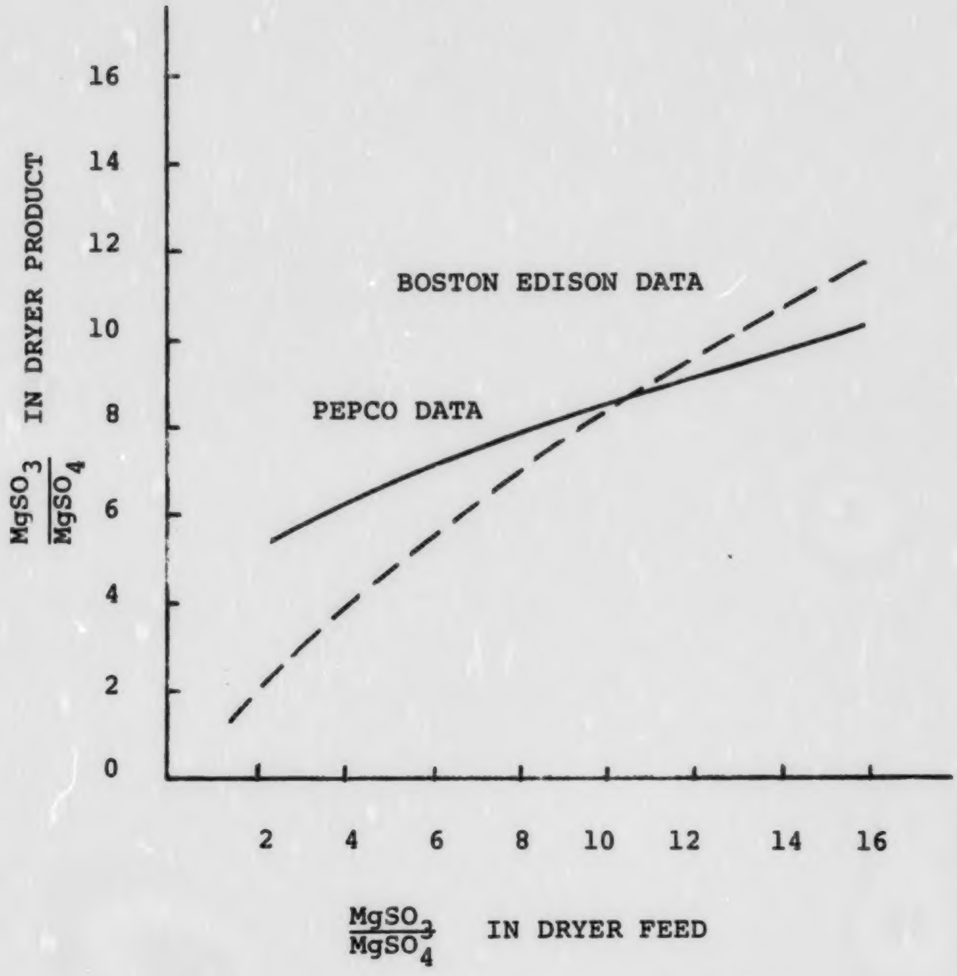
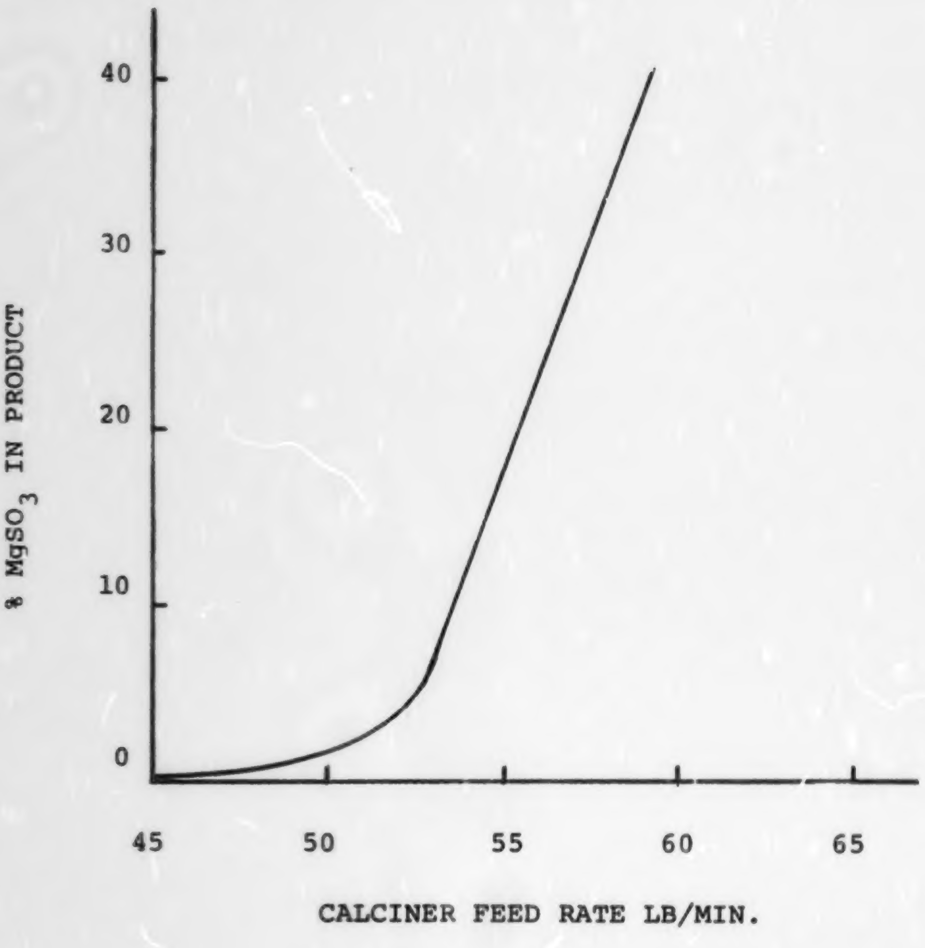


FIGURE 18

CALCINER OPERATIONS ON PEPCO $MgSO_3$

EFFECT OF FEED RATE ON $MgSO_3$ IN PRODUCT



16% which eliminated the slides in the calcination step. The investigations at Dickerson showed there was a significant difference in the size distribution of the product produced from its dryer compared to material produced at Boston as shown in Tables 26 and 27.

Table 26 shows the average values and standard deviations for 25 Boston dryer samples and 22 PEPCO dryer samples covering the following mesh size ranges: +25, -25, to +50 -50 to +100 and -100 to +200. Table 25 presents a statistical analysis of these results.

TABLE 26
SIZE DISTRIBUTION OF DRYER PRODUCTS

Mesh Size Range	BOSTON 25 Samples Average	BOSTON 25 Samples Std. Deviation	PEPCO 22 Samples Average	PEPCO 22 Samples Std. Deviation
+25	58.1	10.2	22.3	11.4
-25, +50	14.9	4.5	29.5	8.5
-50, +100	9.4	4.4	23.5	8.4
-100, +200	4.7	2.0	11.1	4.0

TABLE 27
COMPARISON OF
BOSTON AND PEPCO SIZE RANGE AVERAGES

Mesh Size Range	Significance Level	Difference (Boston-PEPCO)	99% Confidence Limits
+25	99% +	35.8	27.2-44.5
-25, +50	99% +	14.6	9.2-19.9
-50, +100	99% +	14.1	8.8-19.4
-100, +200	99% +	6.4	3.9-8.9

8.0 DATA

8.1 MONTHLY AVERAGE OPERATING CONDITIONS

Operating data are summarized as montly averages for the planned operational testing period. Included in this information are:

Table 28	Operating Conditions for the FGD System
Table 29	Stream Compositions for the FGD System
Table 30	Operating Conditions for the Regeneration Unit
Table 31	Stream Composition for the Regeneration Unit

8.2 DATA LISTING

Data recorded from the PEPCO operations for the period October 15, 1974 through the end of the program in January 1975 are given in Table 32 . Data were recorded at two hour intervals, which is the interval between successive listing in the table for extended runs.

TABLE 28

OPERATING CONDITIONS FOR THE FGD SYSTEM

MONTHLY DATA AVERAGES

VARIABLE	11/74	12/74	1/75
SO_2 (PPM)			
IN	1057.86	1061.95	1272.22
OUT	193.46	207.90	214.89
% REMOV	81.54	80.57	81.95
pH	7.04	7.06	7.11
POWER PLANT RATE (MW)	174.03	153.98	142.82
DIFF. PRESSURES (IN. H_2O)			
TOTAL	7.49	6.77	6.49
MIST	0.23	0.19	0.21
DRYER	0.76	0.69	0.80
TEMPERATURES (F)			
-ABSORBER INLET	109.26	110.88	108.54
MAGNESIA SLURRY	157.02	139.20	130.33
-DRYER GASOUT	274.65	311.03	254.11
PROD.	179.36	200.62	165.30
FLOWS (GPM)			
CIRC.	10152.78	10823.13	11546.30
CENTFD	72.74	83.09	154.61
CENTRIFUGE TORQ.	34.75	40.38	43.94

TABLE 29

STREAM COMPOSITIONS FOR THE FGD SYSTEMS

MONTHLY DATA AVERAGES

VARIABLE	11/74	12/74	1/75
DRYER			
% SOL.	93.63	96.94	91.30
% $MgSO_3$	68.06	68.12	66.45
% $MgSO_4$	8.28	9.08	7.70
% MgO	4.78	6.82	7.78
CENTRIFUGE			
% SOL.	88.86	88.33	88.45
% $MgSO_3$	41.78	42.81	42.19
% $MgSO_4$	4.44	4.86	4.02
% MgO	3.06	3.03	5.25
RECYCLE			
% SOL.	6.98	10.90	8.72
pH	7.17	7.30	7.22
-FILTRATE			
% $MgSO_4$	17.05	18.95	13.13
-CAKE			
% MgO	3.81	3.95	5.79
MOTHER LIQUOR			
% SOL.	2.46	4.80	2.20
-CAKE			
% MgO	6.14	5.68	9.73

TABLE 30

OPERATING CONDITIONS FOR THE REGENERATION UNIT

MONTHLY DATA AVERAGES

VARIABLE	10/74	11/74	12/74	1/75
CALCINER				
-TEMPERATURES (F)				
MDKILN	1260.59	1361.25	1451.97	1494.86
GASOUT	571.75	534.04	544.39	582.89
PRODEX	224.72	160.00	155.22	169.74
-SOLID FLOWS (PPM)				
$MgSO_3$	50.20	51.62	56.61	55.40
COKE	0.43	0.32	0.30	0.31
NEUT. pH LIQ.	5.25	5.83	6.01	6.40
ACID PLANT FEED GAS (%)				
02	5.76	7.34	7.57	7.40

TABLE 31

STREAM COMPOSITIONS FOR THE REGENERATION UNIT

MONTHLY DATA AVERAGES

VARIABLE	10/74	11/74	12/74	1/75
CALCINER FEED				
% H_2O	11.20	11.38	6.00	6.45
% $MgSO_3$	64.08	67.47	63.78	68.42
% $MgSO_4$	7.89	7.25	9.25	8.63
% MgO	5.80	4.94	7.69	6.28
% CARBON	2.66	1.35	1.76	1.32
CALCINER PRODUCT				
% $MgSO_3$	0.45	0.15	0.92	0.46
% $MgSO_4$	0.90	1.29	1.08	3.74
% MgO	89.78	94.62	93.57	93.41
BULK DENSTIY	27.61	22.73	22.41	21.31

TABLE 32

RUN DATES AND DATA LISTINGS

RUN DATES		
STARTING	ENDING	CASE NO.'S
10/15/74	10/18/74	1-31
11/24/74	12/1/74	128-214
12/6/74	12/10/74	218-277
12/14/74	12/23/74	279-384
12/27/74	12/29/74	386-415
1/12/75	1/14/75	421-451

The following identifies the abbreviated words on Pages 126 - 133:

SO_2 IN PPM = Parts Per Million Of SO_2 Into Scrubber

% SO_2 REM = Percent SO_2 Removed In Scrubber

D GAS T = Dry Gas Temperature

D PROD T = Dry Product Temperature

CIRC GPM = Recycle GPM

FAN AMPS = Current required To Drive Fan

CASE NO.	1 SO2INPPM	2 %SO2 REM	3 PH	4 DELTA P	5 D GAS T	6 D PROD T	7 CIRC GPM	8 FAN AMPS
1	MISSING	MISSING	7.3000	4.0000	MISSING	MISSING	10000.0000	190.0000
2	MISSING	MISSING	7.2000	9.0000	MISSING	MISSING	9800.0000	250.0000
3	MISSING	MISSING	7.2000	8.6000	MISSING	MISSING	10000.0000	230.0000
4	MISSING	MISSING	7.0000	9.0000	MISSING	MISSING	9800.0000	230.0000
5	MISSING	MISSING	6.9000	6.0000	240.0000	150.0000	9800.0000	200.0000
6	MISSING	MISSING	7.0000	5.6000	300.0000	145.0000	10000.0000	200.0000
7	MISSING	MISSING	6.8000	5.5000	380.0000	160.0000	10000.0000	200.0000
8	MISSING	MISSING	6.9000	5.0000	330.0000	160.0000	10000.0000	200.0000
9	MISSING	MISSING	6.9000	5.5000	280.0000	170.0000	10000.0000	200.0000
10	MISSING	MISSING	6.9000	5.5000	270.0000	175.0000	10000.0000	200.0000
11	MISSING	MISSING	6.9000	5.5000	240.0000	175.0000	10050.0000	200.0000
12	MISSING	MISSING	6.8000	5.5000	240.0000	165.0000	9900.0000	190.0000
13	MISSING	MISSING	6.9000	5.5000	255.0000	165.0000	9900.0000	200.0000
14	MISSING	MISSING	6.9000	5.5000	220.0000	160.0000	9900.0000	200.0000
15	MISSING	MISSING	6.9000	6.0000	280.0000	170.0000	9900.0000	200.0000
16	MISSING	MISSING	7.1000	7.6000	320.0000	195.0000	9800.0000	200.0000
17	MISSING	MISSING	7.1000	5.5000	275.0000	180.0000	9800.0000	200.0000
18	MISSING	MISSING	6.9000	6.0000	280.0000	180.0000	9800.0000	200.0000
19	MISSING	MISSING	7.0000	5.2000	280.0000	175.0000	9900.0000	200.0000
20	MISSING	MISSING	7.0000	5.5000	280.0000	170.0000	MISSING	200.0000
21	MISSING	MISSING	7.1000	5.5000	360.0000	170.0000	MISSING	200.0000
22	MISSING	MISSING	7.0000	5.5000	250.0000	165.0000	MISSING	200.0000
23	MISSING	MISSING	6.8000	5.5000	260.0000	185.0000	MISSING	200.0000
24	MISSING	MISSING	6.7000	4.0000	220.0000	175.0000	MISSING	190.0000
25	MISSING	MISSING	6.9000	5.5000	230.0000	165.0000	MISSING	200.0000
26	MISSING	MISSING	6.7000	3.5000	225.0000	160.0000	MISSING	190.0000
27	MISSING	MISSING	6.9000	4.0000	225.0000	160.0000	MISSING	MISSING
28	MISSING	MISSING	6.8000	4.5000	225.0000	165.0000	MISSING	190.0000
29	MISSING	MISSING	7.1000	3.5000	235.0000	165.0000	MISSING	180.0000
30	MISSING	MISSING	7.1000	3.5000	235.0000	165.0000	MISSING	180.0000
31	MISSING	MISSING	7.0000	3.3000	220.0000	155.0000	MISSING	MISSING
32	MISSING	MISSING	6.8000	7.0000	MISSING	MISSING	MISSING	210.0000
33	MISSING	MISSING	6.3000	4.0000	MISSING	MISSING	MISSING	180.0000
34	MISSING	MISSING	6.9000	9.0000	210.0000	155.0000	MISSING	225.0000
35	MISSING	MISSING	6.9000	8.0000	220.0000	165.0000	MISSING	225.0000
36	MISSING	MISSING	6.7000	8.6000	250.0000	170.0000	MISSING	220.0000
37	MISSING	MISSING	6.9000	8.6000	240.0000	175.0000	MISSING	220.0000
38	MISSING	MISSING	6.9000	8.6000	235.0000	175.0000	MISSING	220.0000
39	MISSING	MISSING	6.9000	8.6000	235.0000	175.0000	MISSING	220.0000
40	MISSING	MISSING	7.0000	8.6060	220.0000	170.0000	MISSING	220.0000
41	MISSING	MISSING	6.8000	7.0000	210.0000	160.0000	MISSING	220.0000
42	MISSING	MISSING	6.9000	7.0000	210.0000	160.0000	MISSING	220.0000
43	MISSING	MISSING	7.3000	7.0000	210.0000	160.0000	MISSING	220.0000
44	MISSING	MISSING	7.3000	8.5000	225.0000	160.0000	MISSING	220.0000
45	MISSING	MISSING	7.3000	10.5000	225.0000	168.0000	MISSING	240.0000
46	1110.0000	94.5900	7.4000	13.4000	210.0000	177.0000	MISSING	280.0000
47	1260.0000	93.8100	7.3000	13.2000	260.0000	180.0000	MISSING	280.0000
48	1260.0000	93.8100	MISSING	MISSING	MISSING	MISSING	MISSING	150.0000
49	1260.0000	93.8100	MISSING	MISSING	MISSING	MISSING	MISSING	150.0000
50	MISSING	MISSING	7.2000	2.5000	265.0000	MISSING	MISSING	160.0000
51	MISSING	MISSING	6.6000	1.4000	220.0000	165.0000	MISSING	150.0000
52	MISSING	MISSING	7.1000	1.5000	270.0000	170.0000	MISSING	150.0000
53	1260.0000	95.2400	6.9000	3.0000	300.0000	185.0000	MISSING	165.0000
54	1200.0000	84.0000	6.9000	3.0000	300.0000	185.0000	MISSING	175.0000
55	1200.0000	84.0000	6.7000	4.8000	275.0000	180.0000	MISSING	190.0000
56	1260.0000	84.6700	6.8000	2.0000	290.0000	175.0000	MISSING	150.0000
57	MISSING	MISSING	6.3000	6.5000	MISSING	MISSING	12400.0000	220.0000

```
            M  2 %SO2 REM  3 PH         4 DELTA P  5 D GAS T  6 D PROD T  7 CIRC GPM  8 FAN AMPS
E NO.  1 SO2INPF
  58      MISSING      MISSING      6.6000     5.5000      MISSING     MISSING   12400.0000   200.0000
  59      MISSING      MISSING      7.3000     6.2000      MISSING     MISSING   10000.0000   220.0000
  60      MISSING      MISSING      7.3000     6.2000      MISSING     MISSING   10000.0000   200.0000
  61      MISSING      MISSING      6.5000     1.000      260.0000    160.0000   11200.0000   160.0000
  62      MISSING      MISSING      6.8000     3.5000     282.0000    162.0000   11400.0000   190.0000
  63      MISSING      MISSING      6.8000     3.5000     282.0000    162.0000   11400.0000   180.0000
  64      MISSING      MISSING      7.1000     4.0000     282.0000    180.0000   11700.0000   200.0000
  65    1140.0000     77.8900       6.7000     2.5000     310.0000    175.0000   12000.0000   175.0000
  66    1200.0000     79.0000       6.8000     4.2000     280.0000    175.0000   11700.0000   190.0000
  67    1200.0000     79.5000       6.6000     5.0000     265.0000    165.0000   12000.0000   190.0000
  68     870.0000     75.8600       8.0000     6.0000      MISSING     MISSING    7700.0000   220.0000
  69    1080.0000     84.4400       7.2000     6.0000      MISSING     MISSING    7500.0000   220.0000
  70      MISSING      MISSING      6.7000     4.0000     265.0000    150.0000    9500.0000   200.0000
  71      MISSING      MISSING      7.6000     3.0000     280.0000    155.0000    9400.0000   170.0000
  72      MISSING      MISSING      7.6000     3.0000     310.0000    155.0000    9100.0000   180.0000
  73      MISSING      MISSING      7.5000     3.0000     310.0000    155.0000    9300.0000   210.0000
  74     900.0000     83.3300       7.0000     6.5000     295.0000    160.0000   10000.0000   220.0000
  75     900.0000     82.0000       7.1000     5.0000     290.0000    175.0000    9400.0000   200.0000
  76     960.0000     80.6300       6.9000     5.2000     300.0000    185.0000    9400.0000   210.0000
  77     960.0000     77.5000       6.9000     3.0000     270.0000    173.0000    9400.0000   175.0000
  78     900.0000     86.6700       7.0000     7.0000     300.0000    170.0000     MISSING    220.0000
  79     960.0000     83.1300       7.2000     7.0000     270.0000    170.0000     MISSING    220.0000
  80    1020.0000     85.2900       7.3000     8.3000     310.0000    180.0000     MISSING    240.0000
  81    1020.0000     84.1200       7.1000     8.0000     285.0000    182.0000     MISSING    230.0000
  82     960.0000     80.0000       7.3000     MISSING    300.0000    185.0000     MISSING    200.0000
  83     960.0000     84.3800       7.5000     8.0000     260.0000    185.0000     MISSING    235.0000
  84     960.0000     81.2500       7.5000     8.0000     260.0000    185.0000     MISSING    230.0000
  85     960.0000     81.2500       7.5000     8.0000     260.0000    185.0000     MISSING    230.0000
  86    1140.0000     81.0500       7.1000     5.8000     340.0000    190.0000     MISSING    200.0000
  87    1200.0000     78.5000       7.2000     3.0000      MISSING     MISSING     MISSING    170.0000
  88    1260.0000     80.0000       7.0000     3.0000     315.0000    195.0000     MISSING    170.0000
  89     900.0000     76.6700       7.0000     5.2000     315.0000    180.0000     MISSING    200.0000
  90     960.0000     79.3800       7.0000     5.2000     290.0000    170.0000     MISSING    200.0000
  91     900.0000     79.3300       6.9000     7.0000     300.0000    175.0000     MISSING    220.0000
  92     840.0000     80.1200       7.1000     7.0000     310.0000    185.0000     MISSING    220.0000
  93    1080.0000     80.5600       7.0000     7.0000     300.0000    190.0000     MISSING     MISSING
  94     900.0000     84.0000       7.2000     9.4000     280.0000    188.0000     MISSING    250.0000
  95     840.0000     86.4300       7.3000    12.0000     245.0000    182.0000     MISSING    280.0000
  96     900.0000     86.6700       7.2000    13.0000     225.0000    186.0000     MISSING    300.0000
  97     900.0000     86.6700       7.1000    12.5000     280.0000    200.0000     MISSING    300.0000
  98    1080.0000     86.1100       6.8000    11.5000     270.0000    190.0000     MISSING    280.0000
  99    1080.0000     86.1100       6.8000    11.5000     270.0000    190.0000     MISSING    280.0000
 100    1080.0000     86.1100       7.1000     9.5000     250.0000    190.0000     MISSING    260.0000
 101    1020.0000     79.4100       6.8000     MISSING    200.0000    160.0000     MISSING    240.0000
 102     960.0000     79.3800       6.8000     5.0000     240.0000    150.0000     MISSING    180.0000
 103     960.0000     75.0000       7.2000     1.2000     238.0000    155.0000     MISSING    160.0000
 104      MISSING      MISSING      7.3000     7.0000      MISSING     MISSING    8500.0000   240.0000
 105     780.0000     85.3800       7.1000     9.5000      MISSING     MISSING    8200.0000   260.0000
 106    1560.0000      MISSING      6.8000     7.0000     250.0000    175.0000    8600.0000   240.0000
 107    1380.0000     86.9600       7.1000     9.2000     255.0000    175.0000    8600.0000   260.0000
 108    1380.0000     86.9600       7.1000     9.2000      MISSING    180.0000    8600.0000   260.0000
 109    1380.0000     86.9600       7.1000     9.2000      MISSING    180.0000    8600.0000   260.0000
 110    1320.0000     86.3600       7.2000    10.000       MISSING    180.0000    9500.0000   260.0000
 111    1140.0000     84.7400       6.9000    10.000      240.0000    180.0000   11200.0000   260.0000
 112    1140.0000     84.2100       7.1000     8.0000     250.0000    170.0000   11200.0000   230.0000
 113    1150.0000     84.3500       7.1000     8.5000     300.0000    195.0000   11200.0000   230.0000
 114    1080.0000     83.3300       7.1000    10.000      200.0000    180.0000   10400.0000   260.0000
 115    1140.0000     84.2100       7.0000    10.000      200.0000    190.0000   10700.0000   260.0000
 116    1140.0000     84.2100       7.0000    10.000      200.0000    150.0000   11100.0000   260.0000
 117    1140.0000      MISSING      7.0000    10.000      200.0000    150.0000   11100.0000   260.0000
 118    1140.0000     84.2100       7.2000    10.000      200.0000    160.0000   11300.0000   270.0000
```

E NO.	1 SO2INPPM	2 %SO2 REM	3 PH	4 DELTA P	D GAS T	6 D PROD T	7 CIRC GPM	8 FAN AMPS
119	1140.0000	79.4700	7.0000	7.2000	200.0000	155.0000	11400.0000	230.0000
120	1140.0000	79.4700	6.8000	7.2000	210.0000	160.0000	11800.0000	220.0000
121	MISSING	MISSING	7.1000	7.2000	210.0000	160.0000	11800.0000	220.0000
122	MISSING	MISSING	7.3000	8.9000	MISSING	158.0000	12300.0000	230.0000
123	900.0000	81.3300	7.2000	10.2000	212.0000	158.0000	10400.0000	260.0000
124	1200.0000	92.0000	7.1000	11.7000	212.0000	180.0000	10100.0000	280.0000
125	1680.0000	94.6400	7.0000	11.4000	225.0000	177.0000	10400.0000	280.0000
126	1680.0000	94.6400	7.1000	10.000	200.0000	130.0000	10400.0000	250.0000
127	960.0000	81.8800	7.0000	9.6000	220.0000	155.0000	10400.0000	250.0000
128	960.0000	81.8800	7.2000	7.0000	250.0000	130.0000	9200.0000	220.0000
129	900.0000	82.0000	7.1000	7.8000	245.0000	170.0000	9300.0000	230.0000
130	900.0000	85.3300	7.1000	9.7000	242.0000	175.0000	9300.0000	250.0000
131	900.0000	85.2200	7.1000	9.7000	261.0000	183.0000	9300.0000	250.0000
132	900.0000	84.4400	7.0000	10.5000	265.0000	180.0000	9300.0000	250.0000
133	MISSING	MISSING	7.5000	2.5000	MISSING	155.0000	9400.0000	170.0000
134	MISSING	MISSING	7.4000	5.5000	MISSING	160.0000	9500.0000	200.0000
135	900.0000	82.6700	7.0000	6.8000	215.0000	155.0000	9500.0000	220.0000
136	960.0000	82.5000	6.9000	6.8000	203.0000	155.0000	MISSING	220.0000
137	960.0000	85.6300	7.0000	8.6000	230.0000	165.0000	9600.0000	240.0000
138	900.0000	84.0000	MISSING	8.9000	228.0000	165.0000	9700.0000	240.0000
139	900.0000	84.0000	7.1000	8.8000	230.0000	172.0000	9700.0000	240.0000
140	900.0000	84.6700	7.1000	9.0000	220.0000	163.0000	10000.0000	240.0000
141	900.0000	84.6700	7.1000	9.0000	250.0000	172.0000	10100.0000	240.0000
142	900.0000	84.0000	7.0000	9.1000	265.0000	175.0000	10100.0000	240.0000
143	960.0000	84.3800	6.9000	9.1000	272.0000	173.0000	10000.0000	240.0000
144	960.0000	84.3800	7.0000	8.5000	240.0000	160.0000	10000.0000	240.0000
145	900.0000	82.0000	7.1000	8.5000	210.0000	160.0000	10500.0000	240.0000
146	900.0000	82.0000	7.1000	9.3000	220.0000	158.0000	10400.0000	240.0000
147	900.0000	83.3300	7.2000	8.7000	220.0000	153.0000	10100.0000	240.0000
148	960.0000	83.1300	7.3000	9.0000	220.0000	153.0000	10100.0000	240.0000
149	960.0000	82.5000	7.1000	9.0000	240.0000	159.0000	10000.0000	240.0000
150	1020.0000	82.3500	6.9000	9.0000	230.0000	170.0000	10200.0000	240.0000
151	1020.0000	82.3500	6.9000	9.0000	230.0000	170.0000	10200.0000	240.0000
152	1020.0000	82.9400	6.9000	9.0000	230.0000	170.0000	10200.0000	240.0000
153	1020.0000	82.9400	6.9000	9.0000	230.0000	170.0000	10200.0000	240.0000
154	1020.0000	82.9400	6.9000	9.0000	230.0000	170.0000	10200.0000	240.0000
155	1020.0000	81.1800	7.2000	9.1000	270.0000	180.0000	10200.0000	240.0000
156	960.0000	80.6300	7.0000	9.0000	290.0000	180.0000	MISSING	240.0000
157	1080.0000	82.2200	7.0000	8.5000	300.0000	180.0000	MISSING	240.0000
158	1020.0000	81.1800	7.1000	8.5000	280.0000	185.0000	MISSING	240.0000
159	1020.0000	81.1800	7.1000	9.0000	280.0000	190.0000	MISSING	240.0000
160	1020.0000	80.0000	7.0000	9.0000	220.0000	170.0000	MISSING	240.0000
161	1080.0000	80.5600	7.0000	9.0000	220.0000	170.0000	MISSING	240.0000
162	1140.0000	80.5300	7.0000	9.0000	230.0000	180.0000	MISSING	240.0000
163	1140.0000	80.5300	7.0000	9.0000	230.0000	180.0000	MISSING	240.0000
164	1140.0000	79.4700	7.2000	9.0000	225.0000	180.0000	MISSING	240.0000
165	1020.0000	78.2400	7.2000	8.6000	220.0000	190.0000	MISSING	240.0000
166	1080.0000	79.3300	6.9000	9.0000	220.0000	170.0000	MISSING	240.0000
167	1080.0000	78.3300	7.1000	8.7000	220.0000	170.0000	MISSING	240.0000
168	1020.0000	80.0000	7.1000	9.0000	225.0000	170.0000	MISSING	240.0000
169	1020.0000	82.9400	7.0000	9.0000	220.0000	170.0000	MISSING	240.0000
170	1020.0000	82.3500	7.0000	9.0000	220.0000	170.0000	MISSING	240.0000
171	960.0000	81.2500	7.1000	9.0000	220.0000	170.0000	MISSING	240.0000
172	960.0000	78.1300	7.2000	9.0000	220.0000	165.0000	MISSING	240.0000
173	1080.0000	80.0000	6.9000	9.3000	225.0000	175.0000	MISSING	240.0000
174	960.0000	77.2900	7.0000	8.5000	230.0000	170.0000	MISSING	230.0000
175	1140.0000	81.5800	6.9000	10.000	230.0000	170.0000	MISSING	240.0000
176	1200.0000	84.0000	6.9000	9.3000	240.0000	175.0000	MISSING	230.0000
177	1200.0000	84.0000	6.9000	9.3000	240.0000	175.0000	MISSING	220.0000
178	1200.0000	84.0000	6.9000	9.3000	240.0000	175.0000	MISSING	190.0000
179	1200.0000	84.0000	7.0000	4.0000	310.0000	180.0000	MISSING	190.0000

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SE NO.  1 SO2INPPM  2 %SO2 REM  3 PH       4 DELTA P   5 D GAS T  6 D PROD T  7 CIRC GPM  8 FAN AMPS

 180    1200.0000     84.0000     7.0000     4.0000    315.0000    180.0000    MISSING    190.0000
 181    1080.0000     78.8900     6.9000     4.0000    300.0000    165.0000    MISSING    190.0000
 182    1080.0000     78.8900     6.7000     5.1000    400.0000    205.0000    MISSING    180.0000
 183    1080.0000     78.8900     6.9000     3.9000    345.0000    185.0000    MISSING    170.0000
 184    1080.0000     78.3300     6.9000     6.0000    335.0000    185.0000    MISSING    210.0000
 185    1140.0000     82.1100     7.0000     8.0000    325.0000    180.0000    MISSING    230.0000
 186    1080.0000     77.7800     7.1000     5.8000    400.0000    195.0000    MISSING    200.0000
 187    1080.0000     77.7800     6.9000     7.0000    340.0000    160.0000    MISSING    220.0000
 188    1080.0000     77.7800     6.9000     7.0000    340.0000    160.0000    MISSING    190.0000
 189    1140.0000     76.3200     6.7000     6.0000    280.0000    165.0000    MISSING    200.0000
 190    1140.0000     73.6800     6.7000     6.0000    280.0000    165.0000    MISSING    200.0000
 191    1140.0000     73.6800     7.0000     5.8000    365.0000    240.0000    MISSING    200.0000
 192    1140.0000     73.6800     7.0000     5.1000    410.0000    240.0000    MISSING    200.0000
 193    1080.0000     75.5600     7.0000     5.9000    420.0000    253.0000    MISSING    200.0000
 194    1140.0000     77.8900     6.9000     6.2000    420.0000    251.0000    MISSING    200.0000
 195    1080.0000     76.1100     7.0000     6.0000    435.0000    262.0000    MISSING    200.0000
 196    1020.0000     74.7100     7.0000     5.5000    425.0000    256.0000    MISSING    200.0000
 197    1020.0000     74.7100     7.1000     5.5000    450.0000    245.0000    MISSING    200.0000
 198    1140.0000     77.8900     7.0000     6.0000    450.0000    245.0000    MISSING    210.0000
 199    1140.0000     77.3700     7.0000     6.5000    450.0000    260.0000    MISSING    210.0000
 200    1140.0000     76.3200     7.1000     6.4000    430.0000    260.0000    MISSING    210.0000
 201    1140.0000     76.3200     7.0000     7.0000    410.0000    250.0000    MISSING    210.0000
 202    1140.0000     77.3700     7.0000     6.8000    390.0000    240.0000    MISSING    210.0000
 203    1140.0000     76.3200     7.0000     6.6000    365.0000    235.0000    MISSING    210.0000
 204    1140.0000     74.7400     7.1000    MISSING    370.0000    235.0000    MISSING    210.0000
 205    1140.0000     76.3200     7.0000     6.6000    360.0000    230.0000    MISSING    210.0000
 206    1140.0000     76.3200     6.9000     6.5000    360.0000    230.0000    MISSING    210.0000
 207    1140.0000     76.3200     7.0000     6.5000    350.0000    220.0000    MISSING    210.0000
 208    1140.0000     76.3200     7.0000     6.5000    350.0000    210.0000    MISSING    210.0000
 209    1080.0000     75.5600     7.0000     6.5000    350.0000    210.0000    MISSING    210.0000
 210    1080.0000     76.6700     7.1000     6.5000    320.0000    190.0000    MISSING    210.0000
 211    1080.0000     76.6700     7.1000     6.5000    350.0000    190.0000    MISSING    210.0000
 212    1020.0000     74.4700     7.1000     6.6000    340.0000    200.0000    MISSING    210.0000
 213    1020.0000     76.4700     7.1000     6.2000    350.0000    200.0000    MISSING    210.0000
 214     960.0000     72.5000     7.1000     5.4000    375.0000    250.0000    MISSING    210.0000
 215     960.0000     75.6300     7.3000     3.0000     MISSING     MISSING    MISSING    170.0000
 216     960.0000     75.6300     7.3000     3.0000     MISSING     MISSING    MISSING    200.0000
 217    1020.0000     79.4100     7.2000     5.5000    200.0000     MISSING    MISSING    200.0000
 218    1020.0000     76.4700     7.0000     5.2000    250.0000    185.0000    MISSING    200.0000
 219    1020.0000     76.4700     7.0000     5.2000    250.0000    185.0000    MISSING    200.0000
 220    1020.0000     76.4700     7.0000     5.2000    250.0000    185.0000    MISSING    200.0000
 221    1200.0000     MISSING     7.5000     6.6000    310.0000    195.0000    MISSING    210.0000
 222    1020.0000     80.5900     7.1000     8.0000    225.0000    170.0000    MISSING    230.0000
 223    1110.0000     83.7800     7.1000     9.5000    225.0000    180.0000    MISSING    240.0000
 224    1080.0000     77.7800     7.1000     9.5000    225.0000    210.0000    MISSING    240.0000
 225    1080.0000     82.7800     7.1000     9.0000    275.0000    210.0000    MISSING    240.0000
 226    1080.0000     78.3300     7.0000     8.4000    280.0000    210.0000    MISSING    230.0000
 227    1140.0000     83.6800     7.0000     8.4000    275.0000    205.0000    MISSING    240.0000
 228    1080.0000     81.6700     7.1000     8.4000    270.0000    210.0000    MISSING    240.0000
 229    1140.0000     82.6300     7.1000     8.5000    285.0000    210.0000    MISSING    240.0000
 230    1140.0000     76.3200     7.0000     9.0000    270.0000    200.0000    MISSING    240.0000
 231    1140.0000     76.3200     7.1000     9.0000    260.0000    190.0000    MISSING    240.0000
 232    1140.0000     76.3200     7.1000     9.0000    260.0000    190.0000    MISSING    240.0000
 233    1140.0000     76.3200     7.2000     9.0000    250.0000    190.0000    MISSING    240.0000
 234    1140.0000     76.3200     7.2000     9.0000    250.0000    190.0000    MISSING    240.0000
 235    1140.0000     76.3200     7.2000     9.0000    250.0000    190.0000    MISSING    240.0000
 236     840.0000     75.0000     7.2000     7.0000    280.0000    195.0000    MISSING    240.0000
 237    1080.0000     81.6700     7.1000     8.4000    310.0000    210.0000    MISSING    240.0000
 238    1140.0000     78.9500     7.0000     8.5000    280.0000    190.0000    MISSING    240.0000
 239    1140.0000     78.9500     7.1000     8.5000    275.0000    200.0000    MISSING    240.0000
 240    1080.0000     77.7800     7.0000     8.5000    300.0000    210.0000    MISSING    240.0000
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# NO.  1 SO2INPPM  2 %SO2 REM  3 PH        4 DELTA P   5 D GAS T   6 D PROD T  7 CIRC GPM  8 FAN AMPS
  241  1140.0000   82.6300     7.0000      8.5000      340.0000    240.0000      MISSING   240.0000
  242  1140.0000   81.5800     6.8000      8.6000      305.0000    215.0000      MISSING   240.0000
  243  1140.0000   78.9500     7.0000      8.5000      290.0000    205.0000      MISSING   240.0000
  244    MISSING     MISSING   7.0000      8.5000      330.0000    215.0000      MISSING   240.0000
  245  1380.0000   80.4300     6.7000      7.5000      320.0000    210.0000      MISSING   240.0000
  246  1140.0000   74.7400     6.9000      8.4000      275.0000    205.0000      MISSING   240.0000
  247  1140.0000   74.7400     6.9000      7.4000      250.0000    215.0000      MISSING   230.0000
  248  1230.0000   78.7000     7.0000      6.2000      300.0000    210.0000      MISSING   210.0000
  249  1230.0000   78.7000     7.0000      4.0000      400.0000    185.0000      MISSING   200.0000
  250  1080.0000   72.2200     7.0000      4.0000      410.0000    200.0000      MISSING   180.0000
  251  1140.0000   73.6800     7.0000      4.0000      400.0000    210.0000      MISSING   180.0000
  252  1260.0000   76.1900     7.1000      4.0000      390.0000    225.0000      MISSING   180.0000
  253  1140.0000   73.6800     7.1000      4.0000      450.0000    225.0000      MISSING   180.0000
  254  1380.0000   78.2600     7.0000      4.4000      390.0000    225.0000      MISSING   180.0000
  255  1380.0000   78.2600     7.0000      4.4000      300.0000    200.0000      MISSING   180.0000
  256  1380.0000   78.2600     7.1000      4.4000      380.0000    175.0000      MISSING   180.0000
  257  1170.0000   74.3600     7.1000      4.2000      350.0000    220.0000      MISSING   180.0000
  258  1020.0000   75.8800     7.0000      5.0000      325.0000    200.0000      MISSING   180.0000
  259  1020.0000   75.8800     6.9000      4.4000      325.0000    200.0000      MISSING   180.0000
  260  1080.0000   77.2200     7.0000      5.0000      325.0000    200.0000      MISSING   180.0000
  261  1020.0000   74.1200     7.1000      4.5000      320.0000    205.0000      MISSING   190.0000
  262  1020.0000   74.1200     7.0000      4.5000      320.0000    205.0000      MISSING   190.0000
  263  1140.0000   76.3200     6.9000      6.2000      320.0000    205.0000      MISSING   200.0000
  264  1140.0000   76.3200     6.9000      6.2000      330.0000    205.0000      MISSING   200.0000
  265  1200.0000   74.0000     7.1000      6.0000      320.0000    200.0000      MISSING   200.0000
  266  1200.0000   77.0000     7.0000      6.4000      320.0000    210.0000      MISSING   200.0000
  267  1020.0000   70.5900     6.9000      7.0000      320.0000    205.0000      MISSING   200.0000
  268  1320.0000   77.2700     6.9000      7.0000      325.0000    218.0000      MISSING   205.0000
  269  1140.0000   73.6800     7.1000      5.2000      325.0000    224.0000      MISSING   200.0000
  270  1140.0000   73.6800     6.9000      4.7000      325.0000    215.0000      MISSING   190.0000
  271  1140.0000   73.6800     7.2000       1.000      325.0000    215.0000      MISSING   160.0000
  272  1140.0000   73.6800     7.3000      1.2000      325.0000    215.0000      MISSING   160.0000
  273  1080.0000   72.7800     7.3000      2.8000      325.0000    192.0000      MISSING   170.0000
  274  1080.0000   72.2200     6.9000      4.2000      335.0000    195.0000      MISSING   180.0000
  275  1140.0000   73.6800     7.0000      4.0000      352.0000    210.0000      MISSING   180.0000
  276  1140.0000   73.6800     7.0000      5.7000      352.0000    215.0000      MISSING   190.0000
  277  1200.0000   75.0000     7.0000      7.2000      350.0000    220.0000      MISSING   200.0000
  278  1200.0000   75.0000     7.0000      5.8000      350.0000    220.0000      MISSING   205.0000
  279  1200.0000   80.0000     7.2000      5.0000      325.0000    210.0000    10200.0000  190.0000
  280  1200.0000   80.5000     7.2000      5.0000      325.0000    210.0000    10200.0000  190.0000
  281  1200.0000   80.5000     7.2000      5.0000      325.0000    210.0000    10200.0000  190.0000
  282  1080.0000   77.7800     7.2000      5.5000      325.0000    209.0000    10200.0000  190.0000
  283  1020.0000   76.4700     7.2000      4.8000      325.0000    215.0000    10200.0000  190.0000
  284  1080.0000   81.6700     7.2000      6.2000      320.0000    220.0000    10200.0000  200.0000
  285  1080.0000   77.7800     7.2000      6.2000      320.0000    215.0000    10200.0000  200.0000
  286   960.0000   74.3800     7.1000      2.5000      318.0000    225.0000    10200.0000  150.0000
  287  1020.0000   81.7600     6.9000      6.3000      280.0000    190.0000    10200.0000  200.0000
  288   960.0000   81.2500     6.9000      6.0000      277.0000    197.0000    10200.0000  200.0000
  289  1020.0000   84.7100     6.9000      8.0000      270.0000    195.0000    10200.0000  220.0000
  290  1080.0000   83.8900     7.0000      7.0000      265.0000    210.0000    10200.0000  210.0000
  291  1080.0000   83.8900     7.2000      7.0000      265.0000    200.0000    10200.0000  210.0000
  292  1080.0000   83.8900     7.3000      7.0000      265.0000    200.0000    10200.0000  210.0000
  293  1080.0000   83.3300     7.2000      7.0000      265.0000    200.0000    10200.0000  210.0000
  294  1080.0000   83.6100     7.2000      6.5000      270.0000    215.0000    10200.0000  210.0000
  295  1050.0000   84.0000     7.2000      6.7000      265.0000    200.0000    10200.0000  220.0000
  296  1020.0000   82.6500     7.0000      6.7000      265.0000    195.0000    10200.0000  220.0000
  297  1050.0000   82.0000     7.0000      6.7000      280.0000    195.0000    10200.0000  220.0000
  298  1080.0000   80.5600     6.9000      6.5000      280.0000    195.0000    10200.0000  220.0000
  299  1140.0000   84.2100     7.0000      8.2000      280.0000    205.0000    10200.0000  230.0000
  300  1070.0000   82.6500     6.8000      5.0000      280.0000    205.0000    10200.0000  230.0000
  301  1140.0000   83.6800     7.0000      7.0000      285.0000    205.0000    10200.0000  220.0000
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CASE NO.  1 SO2INPPM  2 %SO2 REM  3 PH       4 DEL      5 D GAS T  6 D PROD T  7 CIRC GPM  8 FAN AMPS

     302  1140.0000    81.5800    7.0000    6.7000    280.0000   200.0000  10200.0000   210.0000
     303  1140.0000    83.6800    7.1000    6.5000    280.0000    MISSING  10200.0000   210.0000
     304  1140.0000    82.1100    7.0000    6.0000     MISSING    MISSING  10200.0000   210.0000
     305  1140.0000    82.6300    7.2000    5.5000    285.0000   195.0000  10200.0000   210.0000
     306   930.0000    83.8700    7.3000    6.3000    280.0000   192.0000     MISSING   210.0000
     307   960.0000    82.8100    7.3000    6.0000    280.0000    MISSING  10000.0000    MISSING
     308   840.0000    82.1400    7.3000    4.7000    305.0000   185.0000  10500.0000   150.0000
     309   900.0000    83.3300    7.1000    6.0000    310.0000   185.0000  10800.0000   200.0000
     310   900.0000    85.3300    6.9000    8.2000    310.0000   190.0000  10800.0000   220.0000
     311   900.0000    85.3300    7.0000    8.3000    310.0000   203.0000  10800.0000   220.0000
     312   840.0000    84.2900    7.1000    7.5000    300.0000   205.0000  10400.0000   220.0000
     313   840.0000    85.0000    7.1000    7.0000    300.0000   200.0000  10600.0000   220.0000
     314   840.0000    85.0000    7.2000    7.0000     MISSING    MISSING  10600.0000   220.0000
     315   840.0000    83.5700    7.2000    7.2000    300.0000   200.0000  10600.0000   220.0000
     316   810.0000    77.0400    7.3000    2.3000    300.0000   190.0000     MISSING   160.0000
     317  1080.0000    86.1100    7.1000    8.0000    310.0000   195.0000  11900.0000   180.0000
     318   840.0000    85.0000    7.1000    6.0000    310.0000   205.0000     MISSING   200.0000
     319   810.0000    87.7800    7.0000    9.0000    300.0000   210.0000  11600.0000   250.0000
     320   840.0000    85.7100    7.1000   10.6000    280.0000   200.0000  11600.0000   250.0000
     321   900.0000    85.3300    7.1000   10.4000    250.0000   185.0000  11600.0000   250.0000
     322   960.0000    86.2500    7.0000   12.0000    260.0000   180.0000  11600.0000   250.0000
     323   960.0000    85.6300    7.1000    10.000    260.0000   180.0000  11600.0000   250.0000
     324  1020.0000    84.1200    7.0000    9.8000    260.0000   183.0000  10600.0000   250.0000
     325   960.0000    83.7500    7.1000    10.000    265.0000   180.0000  10800.0000   250.0000
     326   960.0000    83.7500    7.1000    9.8000    267.0000   175.0000  10800.0000   250.0000
     327  1080.0000    84.4400    7.1000    10.000    272.0000   175.0000  10800.0000   250.0000
     328  1100.0000    83.0900    7.0000    9.5000    300.0000   165.0000  10900.0000   230.0000
     329  1140.0000    83.6800    7.0000    9.5000    300.0000   170.0000  10900.0000   230.0000
     330  1140.0000    81.5800    7.1000    7.5000    350.0000   220.0000  11200.0000   210.0000
     331  1140.0000    81.5800    7.1000    9.5000    350.0000   215.0000  11200.0000   230.0000
     332  1200.0000    83.0000    7.0000    9.4000    300.0000   190.0000  11200.0000   230.0000
     333  1200.0000    83.0000    7.0000    9.6000    320.0000   190.0000  11200.0000   230.0000
     334  1200.0000    82.0000    7.0000    8.4000    330.0000   185.0000  11200.0000   210.0000
     335  1140.0000    83.1600    7.0000    8.4000    320.0000   190.0000  11200.0000   210.0000
     336  1140.0000    82.6300    7.0000    8.0000    320.0000   185.0000  10800.0000   220.0000
     337  1080.0000    82.7800    7.0000    8.2000    325.0000   190.0000  11000.0000   220.0000
     338  1080.0000    80.5600    7.0000    8.2000    325.0000   190.0000  11000.0000   220.0000
     339  1080.0000    82.2200    7.0000    8.4000    335.0000   195.0000  11100.0000   220.0000
     340  1080.0000    82.2200    7.0000    8.4000    335.0000   192.0000  11100.0000   220.0000
     341  1200.0000    78.5000    7.0000    2.8000     MISSING    MISSING  11200.0000   170.0000
     342  1260.0000    80.9500    7.1000    6.6000    350.0000   185.0000  11200.0000   200.0000
     343  1320.0000    80.9100    7.1000    6.6000    370.0000   205.0000  11200.0000   200.0000
     344  1380.0000    80.8700    7.0000    7.0000    370.0000   230.0000  11200.0000   200.0000
     345  1200.0000    78.0000    7.0000    3.3000    370.0000   235.0000  10900.0000   160.0000
     346  1140.0000    75.7900    7.1000    3.5000    370.0000   220.0000  10700.0000   160.0000
     347  1140.0000    76.5800    7.0000    3.7000    385.0000   220.0000  10700.0000   160.0000
     348  1110.0000    76.7600    7.1000    3.5000    385.0000   210.0000  10700.0000   160.0000
     349  1140.0000    76.3200    7.0000    3.3000    380.0000   215.0000  11000.0000   160.0000
     350  1080.0000    77.7800    7.0000    3.5000    380.0000   210.0000  11000.0000   160.0000
     351   960.0000    80.6300    7.0000    3.6000    385.0000   160.0000  11000.0000   160.0000
     352   900.0000    80.0000    7.1000    3.6000    380.0000   160.0000  11000.0000   160.0000
     353   960.0000    78.1300    7.1000    3.6000    390.0000   220.0000  11000.0000   160.0000
     354  1020.0000    78.2400    7.0000    3.8000    400.0000   220.0000  11000.0000   160.0000
     355  1020.0000    76.4700    7.0000  [illegible] 400.0000   215.0000  11000.0000   160.0000
     356  1080.0000    77.7800    7.1000  [illegible] 400.0000   210.0000  11000.0000   160.0000
     357   960.0000    78.1300    7.1000    3.0000    400.0000   205.0000  10900.0000   160.0000
     358   990.0000    80.3000    7.1000    4.2000    410.0000   230.0000  10900.0000   180.0000
     359   990.0000    81.2100    7.0000    5.8000    410.0000   230.0000  10900.0000   200.0000
     360  1050.0000    81.7100    7.1000    5.6000    410.0000   235.0000  10900.0000   200.0000
     361  1080.0000    82.2200    7.0000    6.5000    400.0000   240.0000  10900.0000   200.0000
     362  1080.0000    81.1100    7.0000    6.5000    340.0000   215.0000  10900.0000   200.0000
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CASE NO.	1 SO2INPPM	2 %SO2 REM	3 PH	4 DELTA P	5 D GAS T-	6 D PROD T	7 CIRC GPM	8 FAN AMPS
363	1140.0000	80.0000	6.9000	6.5000	335.0000	205.0000	10900.0000	200.0000
364	1140.0000	81.5800	6.8000	6.5000	335.0000	208.0000	10900.0000	200.0000
365	1140.0000	81.0500	7.0000	6.6000	330.0000	210.0000	11200.0000	200.0000
366	1140.0000	81.5800	7.1000	6.6000	330.0000	215.0000	11200.0000	200.0000
367	1140.0000	81.5800	7.1000	6.8000	330.0000	205.0000	11200.0000	200.0000
368	1140.0000	80.5300	7.0000	6.8000	330.0000	210.0000	11200.0000	200.0000
369	1140.0000	81.3200	7.2000	6.2000	330.0000	210.0000	11200.0000	200.0000
370	1110.0000	81.8900	7.1000	6.5000	335.0000	215.0000	11000.0000	200.0000
371	1110.0000	82.1600	7.2000	6.5000	335.0000	210.0000	11000.0000	200.0000
372	1080.0000	82.2200	7.1000	6.4000	335.0000	210.0000	11000.0000	200.0000
373	1080.0000	82.2200	7.0000	6.3000	335.0000	205.0000	11000.0000	200.0000
374	840.0000	82.8600	7.2000	6.3000	335.0000	180.0000	11000.0000	200.0000
375	900.0000	83.3300	7.1000	7.0000	365.0000	197.0000	11000.0000	210.0000
376	900.0000	83.3300	7.1000	7.0000	365.0000	197.0000	11000.0000	210.0000
377	900.0000	83.3300	7.1000	7.0000	365.0000	197.0000	11000.0000	210.0000
378	900.0000	83.3300	7.1000	7.0000	365.0000	197.0000	11000.0000	210.0000
379	900.0000	83.3300	7.1000	7.0000	365.0000	197.0000	11000.0000	210.0000
380	900.0000	83.3300	6.9000	8.0000	360.0000	210.0000	10900.0000	220.0000
381	900.0000	83.3300	6.9000	8.5000	350.0000	220.0000	10800.0000	220.0000
382	840.0000	83.5700	7.0000	8.5000	350.0000	230.0000	10800.0000	220.0000
383	840.0000	84.2900	7.0000	8.8000	340.0000	225.0000	10800.0000	220.0000
384	960.0000	80.0000	7.0000	8.5000	330.0000	210.0000	10800.0000	220.0000
385	1080.0000	77.2200	7.0000	4.0000	MISSING	MISSING	9900.0000	200.0000
386	1080.0000	77.2200	7.1000	2.2000	MISSING	MISSING	11000.0000	160.0000
387	780.0000	78.4600	7.2000	5.9000	MISSING	MISSING	10900.0000	200.0000
388	780.0000	78.4600	7.2000	5.9000	MISSING	MISSING	10900.0000	200.0000
389	900.0000	81.3300	7.2000	5.9000	260.0000	185.0000	10900.0000	200.0000
390	900.0000	82.6700	7.2000	5.9000	260.0000	185.0000	10900.0000	210.0000
391	900.0000	88.0000	7.2000	5.9000	260.0000	185.0000	10900.0000	220.0000
392	900.0000	88.0000	7.2000	8.5000	260.0000	185.0000	10900.0000	240.0000
393	960.0000	87.5000	7.1000	8.8000	265.0000	185.0000	11000.0000	240.0000
394	900.0000	86.6700	7.1000	9.2000	265.0000	185.0000	11000.0000	260.0000
395	900.0000	86.6700	7.1000	9.2000	260.0000	175.0000	11000.0000	260.0000
396	900.0000	86.6700	7.1000	9.2000	260.0000	175.0000	11000.0000	260.0000
397	1020.0000	88.2400	7.1000	9.2000	255.0000	180.0000	11000.0000	260.0000
398	900.0000	86.6700	7.1000	9.4000	280.0000	170.0000	10850.0000	260.0000
399	900.0000	87.3300	7.0000	10.3000	285.0000	192.0000	10850.0000	260.0000
400	900.0000	87.3300	7.0000	10.3000	285.0000	192.0000	10850.0000	260.0000
401	1000.0000	85.6000	7.0000	9.0000	285.0000	192.0000	10850.0000	260.0000
402	1140.0000	86.8400	7.0000	9.0000	210.0000	170.0000	10850.0000	260.0000
403	1020.0000	87.6500	7.0000	10.000	210.0000	170.0000	10850.0000	260.0000
404	1020.0000	88.2400	7.0000	10.000	210.0000	170.0000	10850.0000	260.0000
405	1020.0000	88.2400	7.1000	10.000	200.0000	170.0000	11200.0000	260.0000
406	960.0000	85.0000	7.1000	10.000	200.0000	170.0000	11200.0000	260.0000
407	960.0000	85.0000	7.1000	10.000	200.0000	170.0000	11200.0000	260.0000
408	960.0000	86.2500	7.2000	10.000	200.0000	170.0000	11200.0000	260.0000
409	1080.0000	86.1100	7.1000	10.2000	MISSING	175.0000	10700.0000	260.0000
410	1040.0000	86.7300	7.1000	10.000	275.0000	175.0000	11050.0000	260.0000
411	1080.0000	87.2200	7.0000	10.000	275.0000	200.0000	11050.0000	260.0000
412	1140.0000	87.3700	7.0000	10.000	275.0000	200.0000	11050.0000	260.0000
413	900.0000	86.6700	7.2000	10.000	275.0000	200.0000	11100.0000	260.0000
414	1380.0000	78.7000	7.0000	3.5000	250.0000	180.0000	11400.0000	180.0000
415	1500.0000	80.4000	7.0000	3.0000	260.0000	180.0000	11400.0000	180.0000
416	1060.0000	81.3200	7.0000	5.0000	MISSING	MISSING	11400.0000	220.0000
417	1020.0000	81.1800	7.0000	5.0000	MISSING	MISSING	11400.0000	220.0000
418	1440.0000	85.8300	7.0000	5.5000	MISSING	MISSING	MISSING	220.0000
419	1440.0000	85.8300	7.0000	6.0000	260.0000	170.0000	MISSING	230.0000
420	1200.0000	82.5000	7.1000	7.0000	260.0000	180.0000	MISSING	220.0000
421	1140.0000	81.5800	6.9000	7.0000	260.0000	180.0000	11200.0000	230.0000
422	1200.0000	83.5000	7.1000	7.4000	260.0000	180.0000	11700.0000	230.0000
	1140.0000	80.5300	7.2000	6.3000	2 0000	170.0000	11700.0000	210.0000

```
CASE NO.  1 SO2INPPM  2 %SO2 REM  3 PH      4 DEL   P    5 D GAS T   6 D PROD T  7 CIRC GPM   8 FAN AMPS
   424    1140.0000     79.4700     7.1000      6.5000    260.0000    170.0000  11900.0000    210.0000
   425    1440.0000     82.0800     6.9000      6.2000    265.0000    170.0000  11600.0000    210.0000
   426    1560.0000     82.3100     7.0000      6.1000    265.0000    190.0000  11600.0000    210.0000
   427    1560.0000     81.5400     7.1000      6.0000    260.0000    195.0000  11500.0000    210.0000
   428    1560.0000     MISSING     7.2000      6.0000    263.0000    198.0000  11300.0000    210.0000
   429    1560.0000     MISSING     7.1000      6.0000    260.0000    190.0000  11400.0000    210.0000
   430    1560.0000     MISSING     7.2000      6.0000    260.0000    188.0000  11400.0000    210.0000
   431    1560.0000     MISSING     7.1000      6.0000    260.0000    187.0000  11400.0000    210.0000
   432    1500.0000     MISSING     7.1000      6.2000    265.0000    190.0000  11400.0000    210.0000
   433    1440.0000     MISSING     7.2000      6.2000    260.0000    192.0000  11300.0000    210.0000
   434    1440.0000     MISSING     7.2000      6.2000    250.0000    190.0000  11300.0000    210.0000
   435    1320.0000     MISSING     7.1000      6.4000    260.0000    190.0000  11300.0000    210.0000
   436    1500.0000     MISSING     7.2000      6.0000    260.0000    190.0000  11600.0000    210.0000
   437     870.0000     68.2800     7.2000      6.0000    260.0000    145.0000  11000.0000    210.0000
   438     870.0000     68.2800     7.1000      6.3000    270.0000    180.0000  11400.0000    210.0000
   439    1020.0000     76.4700     7.2000      6.3000    270.0000    180.0000  11400.0000    210.0000
   440    1020.0000     77.0600     7.0000      6.8000    285.0000    165.0000  12000.0000    210.0000
   441    1060.0000     76.7900     7.0000      6.8000    300.0000    175.0000  11800.0000    210.0000
   442    1080.0000     76.6700     7.0000      6.9000    310.0000    170.0000  11800.0000    210.0000
   443    1020.0000     74.7100     7.0000      6.9000    305.0000    170.0000  11800.0000    210.0000
   444    1080.0000     75.0000     7.0000      6.9000    305.0000    170.0000  11800.0000    210.0000
   445    1140.0000     75.2600     7.0000      6.9000    300.0000    170.0000  11600.0000    210.0000
   446    1080.0000     73.3300     7.1000      7.0000    325.0000    165.0000  11800.0000    210.0000
   447    1200.0000     MISSING     7.1000      6.9000    300.0000    174.0000  11800.0000    210.0000
   448    1200.0000     MISSING     7.0000      7.0000    310.0000    175.0000  11800.0000    210.0000
   449    1200.0000     MISSING     7.0000      7.0000    310.0000    175.0000  11800.0000    210.0000
   450    1200.0000     MISSING     7.1500      6.5000    305.0000    175.0000  11500.0000    210.0000
   451      MISSING     MISSING     7.2000      6.5000    305.0000    175.0000  11500.0000    210.0000
   452    1560.0000     MISSING     7.1000      3.4000     MISSING     MISSING  11700.0000    190.0000
   453    1560.0000     MISSING     7.1000      3.4000     MISSING     MISSING  11700.0000    170.0000
   454    1050.0000     MISSING     6.9000      1.2000     MISSING     MISSING  11600.0000    160.0000
   455    1050.0000     MISSING     6.9000     MISSING     MISSING     MISSING  11600.0000    170.0000
   456    1500.0000     82.0000     7.4000      4.5000     MISSING     MISSING  11600.0000    200.0000
   457    1500.0000     84.0000     7.2000      4.0000    200.0000    125.0000  11600.0000    200.0000
   458    1500.0000     84.0000     7.1000      4.3000    200.0000    120.0000  11600.0000    200.0000
   459    1500.0000     84.0000     7.1000     MISSING    200.0000    120.0000  11600.0000    200.0000
   460    1500.0000     84.4000     7.1000      4.4000    200.0000    125.0000  11200.0000    200.0000
   461    1380.0000     89.1300     7.2000      7.8000    200.0000    130.0000  11200.0000    250.0000
   462    1140.0000     88.4200     6.9000      8.0000    210.0000    155.0000  11300.0000    250.0000
   463      MISSING     MISSING     7.1000      7.5000    200.0000     MISSING  11300.0000    240.0000
   464    1140.0000     89.4700     7.2000      9.0000    210.0000    150.0000  11700.0000    260.0000
   465    1080.0000     89.4400     7.1000      9.2000    200.0000    150.0000     MISSING    250.0000
   466    1060.0000     89.0600     MISSING     9.3000    170.0000    145.0000     MISSING     MISSING
   467    1320.0000     88.6400     7.2000      9.4000    200.0000    135.0000  11700.0000    240.0000
   468    1320.0000     89.0900     7.1000      9.0000    200.0000    140.0000  11700.0000    260.0000
   469    1320.0000     89.5500     7.1000      9.0000    200.0000    135.0000  11700.0000    260.0000
   470    1260.0000     89.5200     7.1000      8.8000    200.0000    140.0000  11700.0000    260.0000
   471      MISSING     MISSING     7.5000      1.2000     MISSING     MISSING  11400.0000    160.0000
   472    1140.0000     MISSING     7.4000     10.8000     MISSING     MISSING  11400.0000    240.0000
   473      MISSING     MISSING     7.4000      8.0000     MISSING    115.0000  11400.0000    240.0000

NUMBER OF CASES READ. . . . . . . . . . . . . .     473     VARIABLE                          STANDARD   TOTAL
                                                            NO. NAME               MEAN      DEVIATION  FREQUENCY

                                                             1 SO2INPPM          1092.036     163.471     403
                                                             2 %SO2 REM            81.224       4.444     382
                                                             3 PH                   7.050       0.166     469
                                                             4 DELTA P              6.873       2.290     466
                                                             5 D GAS T            287.215      57.256     429
                                                             6 D PROD T           186.475      24.718     432
                                                             7 CIRC GPM         10725.461     911.543     278
                                                             8 FAN AMPS           215.052      28.681     468
                                                             9 C TORQUE            37.770       8.549     437
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10.0 CONVERSION FROM ENGLISH TO METRIC UNITS

TO CONVERT FROM	TO	MULTIPLY BY
Atmosphere (normal)	Bar	1.01325
Atmosphere (normal)	Pascal	101,325
Barrel (42 US gallons)	Meter^3	0.15899
British thermal unit (BTU)	Joule	1055.1
BTU/Hour	Watt	0.29307
BTU/pound mass	Joules/gram	2.32600
BTU/pound mass - °F	Joules/gram - °K	4.18680
Foot	Meter	0.30480
Foot^2	Meter^2	0.09290
Foot^3	Meter^3	0.02832
Foot^3/minute	Meter^3/minute	0.02832
Foot-pound force	Joule	1.35582
Gallon (US)	Meter^3	0.00379
Gallon (US)/minute	Meter^3/hour	0.22712
Grain	Milligram	64.7989
Horsepower	Kilowatt	0.74570
Inch	Centimeter	2.5400
Inch H_2O (60°F)	Kilopascal	0.24884
Pound Force	Newton	4.44822
Pound mass av	Kilogram	0.45359
Pound mass av	Metric ton (tonne)	0.0004536
Pound force/inch^2	Kilopascal	6.89476
Pound mass/foot^3	Kilograms/meter^3	16.0185
°Rankine	°Kelvin	0.55556
Ton mass (US short)	Kilogram	907.185
Ton mass (US long)	Kilogram	1016.05

APPENDIX I

POST OPERATION INSPECTION REPORT

DATE OF INSPECTION - Jan. 17-24, 1975

(NOTE: Refer to Figure I-1, where numbers refer to following sections)

Plumb Bob Control Mechanism

Operation has been satisfactory.

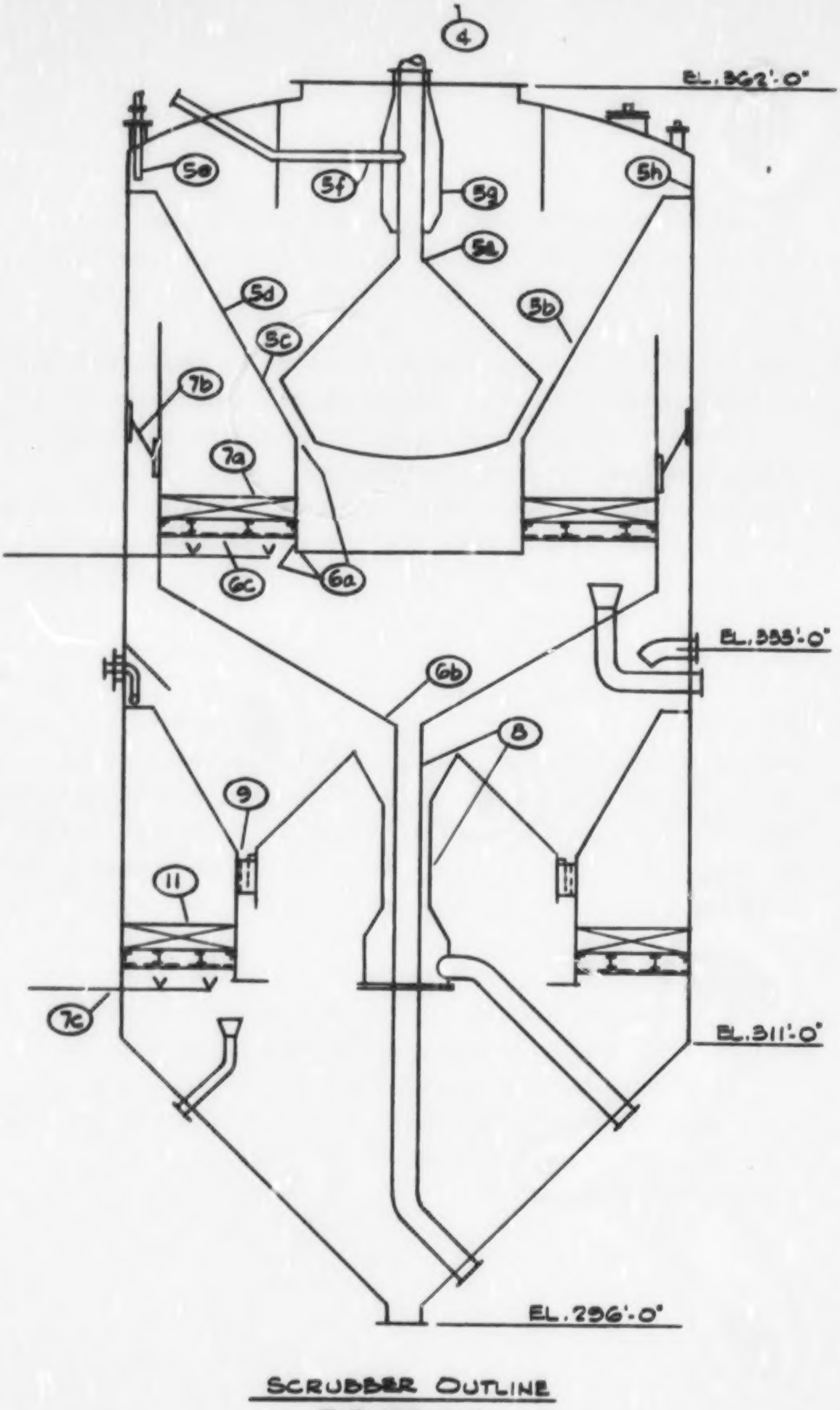
Upper Section of 1st Stage (5)

(a) The polyester coating was gone from the stem and the top 8" of sloping sides of the plumb bob. The metal is type 316L stainless steel at this point and measurements indicated no loss of thickness.

(b) The coating was also gone from most of the 316 stainless steel lip projecting from the sloping wall just above the throat.

(c) In the throat area there were about 10 spots ranging in size from 12 square inches to 30 square inches where the entire coating had been removed from the metal. The metal is 316L stainless steel at this point. It is possible that these coating failures were caused, by debris from the 4 inch stainless steel nozzles which had failed.

(d) Above the projecting lip on the sloping outer wall, there were a number of holes in the lining where the carbon steel had been attacked so that only the lining on the other side of the steel plate prevented a hole in the vessel. The figure is typical of this type of failure.


4
EL. 362'-0"
5e
5f
5g
5h
5a
5d
5b
5c
7b
7a
6c
6a
EL. 333'-0"
6b
8
9
11
7c
EL. 311'-0"
EL. 296'-0"

SCRUBBER OUTLINE

FIGURE I-1

(e) As was discussed previously in the section on corrosion of stainless steel all of the 4-inch tangential nozzles were subjected to extensive failure. Two were still in place, but were in shreds.

(f) The three six-inch stainless steel lines that feed the center manifold were subjected to the same type of attack as the four-inch tangential nozzles and pipes. There was less severity because of the protective coating of brick cast which had been applied over these pipes. Repairs have been carried out by removing the brick cast and coating these six-inch pipes with glass reinforced polyester plastic.

(g) The center manifold was in good condition both internally and externally.

(h) There were two holes through the external wall of the vessel adjacent to tangential inlet nozzles. It is not known the extent to which these were caused by normal erosion or by impaction by partially failed inlet nozzles. Except for these two failures the coating at this portion of the vessel was in good condition.

Lower Section of 1st Stage (6)

(a) The polyester lining was in fair condition, with repairs needed at several points in the throat section, at the bottom of the skirt and on the outer surface of the inner skirt.

(b) There was one small (3/8 inch diameter) hole in the bottom of the cone near the vortex breaker where a repair had been made.

(c) There was considerable loss of metal by corrosion of type 316 stainless steel from the hanger rods, clips, clamps and bolts. As indicated previously, corrosion was most severe in areas which had been stressed and the replacement pieces have been stress-relieved before installation.

1st Stage Mist Eliminator Section (7)

(a) The mist eliminators were clean and in excellent condition. However, all of the J hooks and clips used as hold-downs had failed and most were completely missing. These are among the sources of metal that have contributed to failures of rubber-lined pipe. The 2 inch x 0.2 inch straps used as auxiliary hold-downs for the mist eliminators have performed their function well. There is evidence of slight pitting, but no loss of thickness.

(b) The new clevis type 1st stage cone support hangers showed slight signs of pitting, but are structurally in excellent condition. These were of 316 stainless steel and were installed in July 1974. The Inconel 625 pins installed with them were in excellent condition and showed no sign of attack.

(c) The 1st stage mist eliminator wash system has operated satisfactorily.

Upper Section of 2nd Stage (8)

This portion of the scrubber was found to be in generally good condition. There was a small hole (possibly 1/16 inch in diameter) in the stainless steel pipe just below the base.

Inspection Report (Information of two-stage scrubber absorber and the recycle pipe have been presented in the previous report).

2nd Stage Throat Area (9)

Failure of the coating in the area of the 2nd stage throat was very extensive as illustrated in photographs 10-7 and 10-10. Assurance of adequate structural integrity was provided by a series of thickness measurements.

The throat restrictor provided substantial protection for the inner surface of the throat except where there apparently were gaps between the restrictor and the inner surface. Planned repairs and changes include division into an inner and outer ring, a reduction in the total width (to decrease pressure drop), and caulking between the restrictor plates and walls to eliminate flow and erosion at these points.

Gas Systems

(1) Fan

The fan has given excellent service throughout the entire history of the scrubber. Inspection indicated that the rotor, which is of high alloy stainless steel is in a condition apparently as good as new. The rubber lining of the fan housing is also in a condition as good as new. The only problem was a failure in one of the pipes which feeds water to the fan rotor. This pipe had failed presumably as a result of continued vibration. Repairs were effected by welding.

(2) Secondary Mist Eliminator

The secondary mist eliminator was clean and in excellent condition except for some evidences of corrosion of the various 316 stainless steel hanger rods, etc. This was similar to the corrosion observed in the 1st stage of the scrubber although substantially less in extent.

(3) Ducts and Dampers

There was a fair amount of build-up of fly ash in the ducts as would be expected, but the ducts were otherwise in good condition. The major problem with the dampers has been leakage during shut down, permitting a flow of gas backwards through the ducts and scrubber and a progressive increase in fly ash build-up and decrease in damper effectiveness. At the conclusion of operations in late January, whenever the scrubber was shut down there was a very high flow of gas backwards through the scrubber.

A seal ring has now been installed on the outlet damper, with the expectation of drastically reducing (and possibly eliminating) this problem. It is also expected that better continuity of operation will be obtained and this in itself will permit better performance of the dampers on the few occasions when shut downs are expected to be necessary.

Liquid Systems

(1) 1st Stage Recycle Pumps

These have performed excellently at all times. Inspection through the suction piping indicates an appearance "like new."

(2) Thickeners

These have performed satisfactorily. The south thickener had a number of failures in the coating which were repaired. The north thickener, in contrast, had no coating failures. The bolts that hold the arms of the thickener rakes were corroded excessively in the south thickener, but little or no corrosion was observed in the similar bolts in the north thickener.

(3) Thickener Underflow System

This system has performed satisfactorily and inspection did not indicate any defects.

(4) Thickener Overflow Return System

This system has performed satisfactorily from a mechanical standpoint and inspection did not indicate any defects.

(5) 2nd Stage Recycle Pumps

These pumps have given generally satisfactory performance. The impeller on each of the three pumps has been replaced once. In a commercial installation it is expected that rubber covered impellers would be specified and that impeller replacement would not be necessary. At the conclusion of the operating period in late January 1975, performance curves were obtained on the three pumps as indicated in Figure I-2.

(6) Mother Liquor Tank

This tank has given satisfactory performance. The gears on the agitator were replaced on one occasion. Presumably their failure was due to a build up of solids in the bottom of the tank leading to excessive side thrusts on the agitator.

(7) Mother Liquor Pumps

Operation has been satisfactory.

(8) MgO Premix Tank

Operation of this tank has been satisfactory since the changes were made in the summer of 1974.

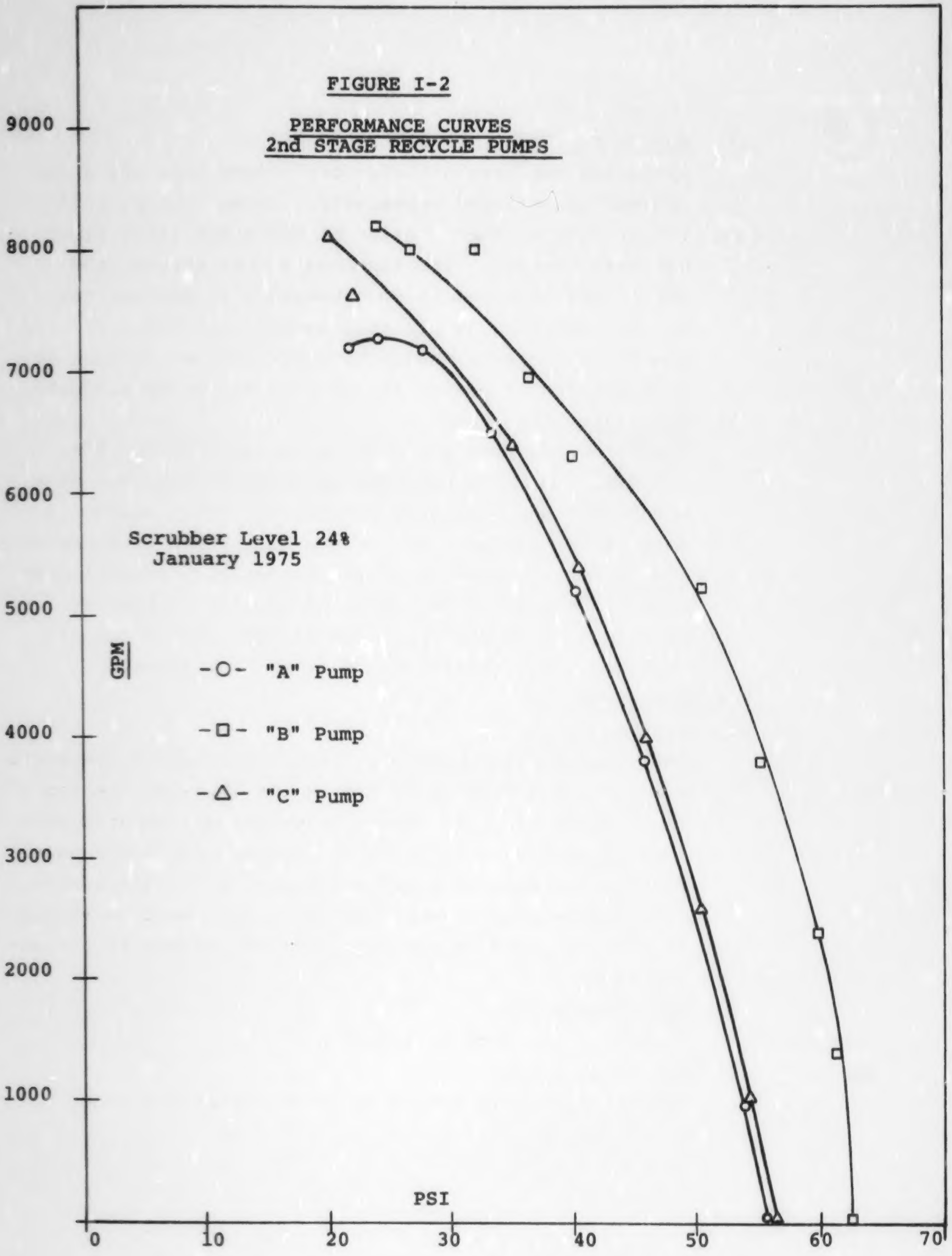


FIGURE I-2

PERFORMANCE CURVES
2nd STAGE RECYCLE PUMPS

(9) MgO Slurry Tank

Operation has been satisfactory except that the steam sparger has plugged repeatedly. During the current boiler outage, the location of the steam inlet line has been changed. The line now enters through the top of the tank and is so arranged that the sparger may be unplugged if plugging should continue. It will also be possible to change to other sparger designs if such becomes necessary in order to eliminate plugging.

(10) MgO Slurry Feed Pumps

Operation has been satisfactory since October 1974, at which time a larger pump was installed as the east MgO pump. One remaining problem with this system is that it is necessary to operate both pumps in order to obtain correct reading of the temperature control element. A change in the location of this element is planned so that the operation may be carried out with only one pump, reserving the other as a spare.

Solids Systems

(1) MgO Unloading

This system has operated satisfactorily in all respects except that a leak developed in the elbow at the top of the loading line, as was to be expected. Repairs were made by welding which will be adequate for the remaining three months of expected operation of this system. For a commercial installation materials more resistant to erosion would be specified for the elbows in unloading lines.

(2) MgO Storage Bin

Operation has been satisfactory.

(3) MgO Weigh Feeder

Operation of this system has been acceptable since changes were made in the summer of 1974. However, the reliability of this system leaves something to be desired and for future commerical installations, other designs would be investigated. There have been several belt failures during the operation of this system. Near the end of operation in late January, the weigh feeder became most unreliable in its operation and during the outage, moderately extensive repairs and replacements have been carried out.

(4) Centrifuge

Operation of the centrifuge has been generally quite satisfactory and very substantially improved over operation experienced at Boston Edison Company. A major difference between the two installations, of course, is that at Boston Edison the centrifuge feed contained primarily magnesium sulfite trihydrate whereas at PEPCO the feed has been primarily magnesium sulfite hexahydrate. There are four areas that require comment:

(a) There has been no evidence of the type of binding between the conveyor and the bowl (internally) that was so troublesome at Boston Edison. Rotation of the centrifuge was continued most of the time, although there were substantial periods when the machine was shut down and binding might have taken place. The washing procedures developed at Boston Edison were applied at PEPCO from the start of the operations. The extent of internal wear, which is probably directly related to binding, could not be determined. The desirability of dismantling the centrifuge for inspection was recognized, but when weighed against the cost involved, the decision was made to resume operations without inspection of the internal portions of the centrifuge.

(b) There was excessive erosion-corrosion of the case at the liquid end, as shown in photograph 3-2. A patch was applied around the entire 180° circumference of the casing. Wear continued and the casing, which was originally 1/4 inch thick, was worn through completely, exposing the patch for a longitudinal distance of about 1-1/2 inches. Although replacement patching internally with stainless steel would probably have provided a permanent cure to this erosion problem, the decision was made, in the interest of economy, to patch the casing with carbon steel. A coupon of 316 stainless steel has been installed and observation will be made at the conclusion of the operating period as to the ability of type 316 to withstand this service.

(c) One plow failed and all of the plows were replaced in September 1974. One plow from the new set failed in January 1975 and a complete third set of plows has now been installed. An improved design of plow may be expected to eliminate this problem in future installations. The photographs 3-5 (the failed plow) and 3-11 (normal plows) suggest that some machines are designed with three bolts per plow.

(5) Screw Conveyors

These have given generally satisfactory performance. There is, of course, considerable evidence of wear, but it is not regarded as excessive. The bearing at the inboard end of the dryer feed screw was dislodged due to a spreading of the bearing support as illustrated in photograph 4-3. The screw continued to operate without any inboard bearing, using the solids in the trough as support, for several weeks. This bearing has now been replaced.

(6) Dryer

The dryer has given excellent performance. Buildup has been limited in amount and it has been possible to clean the dryer usually by an adjustment in operating conditions, although occasionally it has been necessary to use a pipe or rod for this purpose. At the conclusion of operations in late January, the dryer was remarkably clean, as illustrated in photograph 4-1. The refractory of the burner block and the metal casing around it failed as indicated in photograph 4-6 and have been replaced.

(7) Dryer Off-Gas Cyclone

The double gate at the base of this cyclone has never operated properly and a rotary valve has now been installed to replace it. In the past, failure to remove solids from the base of the cyclone has led to an excessive carry-over of fines and recycle to the scrubber. This situation should be substantially improved as a result of the installation of the rotary dust valve.

(8) Dryer Off-Gas Fan

Operation has been satisfactory.

(9) Bucket Elevator

It appears that the bucket elevator was sized for a product of higher density and possibly sized in expectation of continuous operation at a uniform rate. In any event, the capacity of the bucket elevator has been inadequate to handle the full load of the scrubber system and has made it necessary to operate at rates approximating 75% of gas flow capacity and somewhat less than 75% of magnesium sulfite capacity. Larger buckets have now been installed on the bucket elevator and it is expected that capacity operation can be achieved at this point in the system as well as elsewhere.

(10) Magnesium Sulfite Storage Silo

Operation has been satisfactory.

(11) Magnesium Sulfite Loading System

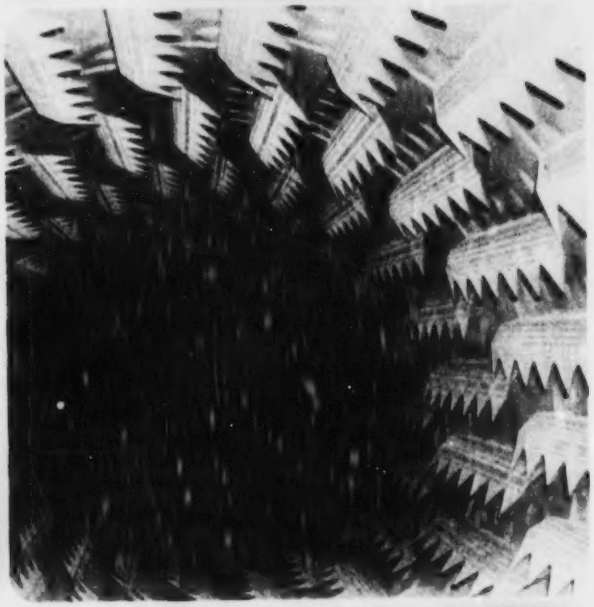
Mechanical operation of this system is considered satisfactory. The repeated problems that have been experienced with excessive spillage of magnesium sulfite during loading are considered due to operating problems rather than to defects in the mechanical design or condition of the equipment.

3-2

3-5

3-11

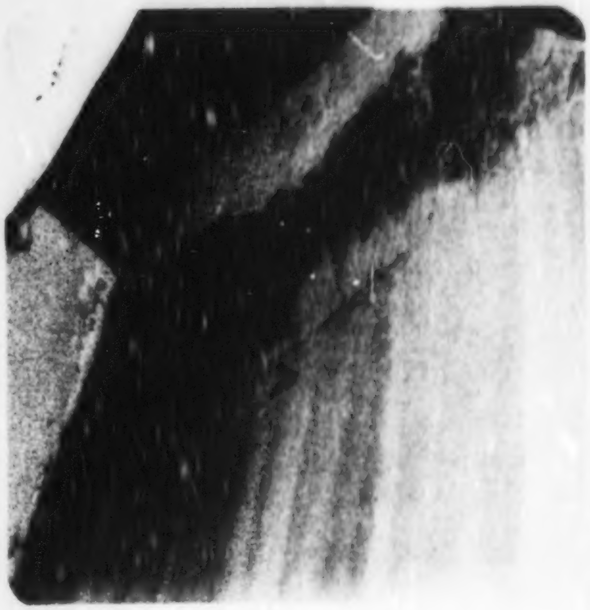
4-1



4-3

4-6

10-7



10-10

APPENDIX 2

SULFUR OXIDE REMOVAL FROM POWER PLANT STACK GAS

Magnesia Scrubbing-Regeneration: Production of Sulfuric Acid

The Thermal Dehydration of Magnesium Sulfite Hexahydrate ($MgSO_3 \cdot 6H_2O$) and Magnesium Sulfite Trihydrate ($MgSO_3 \cdot 3H_2O$);

A mechanistic Study using Thermo-analytical Techniques and the Development of an Analytical Method for Quantitating Mixtures of the Hydrates

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ABSTRACT

The thermal dehydration of $MgSO_3.3H_2O$ and $MgSO_3.6H_2O$ have been studied by differential thermal analysis, differential scanning calorimetry and thermal gravimetric analysis.

Similar studies by other groups led to contradictory conclusions. In this investigation, these results were reconciled and it was concluded that $MgSO_3.6H_2O$, under equilibrium conditions, dehydrates in two steps through the intermediate formation of $MgSO_3.3H_2O$.

One consequence of this work is the development of a simple analytical procedure, using TGA, for the quantitative determination of both hydrates in mixtures and in the presence of thermally inactive material.

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INTRODUCTION

The thermal dehydration of the two known stable hydrates of magnesium sulfite, namely, magnesium sulfite hexahydrate ($MgSO_3 \cdot 6H_2O$) and magnesium sulfite trihydrate ($MgSO_3 . 3H_2O$) have been studied. On the one hand, this work is of practical importance because in the Magnesia Scrubbing Regeneration Process (CHEMICO)(8), the thermal dehydration of the magnesium sulfite hydrates is a significant step. On the other hand, this study is of scientific interest because the mechanism of the dehydration has been the subject of controversy.

In the Magnesia Scrubbing Regeneration Process (CHEMICO), sulfur dioxide in the flue gas is removed by scrubbing with a magnesia (MgO) base slurry. Hydrates of $MgSO_3$ are formed; also $MgSO_4$ is formed by oxidation. A portion of the slurry is withdrawn and centrifuged. The solid is separated from the liquid, which is recycled. The solids are then dehydrated in a dryer and, afterwards, calcined at 1400°F-1600°F with coke added to the solids. Heating is sufficient to effect the thermal decomposition of $MgSO_3$; the coke reduces $MgSO_4$ to $MgSO_3$. The useful products of calcination are MgO, which is recycled back to the slurry, and SO_2, which is converted to sulfuric acid.

Turning to the scientific relevance of this study, thermal dehydration studies have been reported by three groups and the results were apparently contradictory. Okabe and

Hori(5) used three different techniques to study the dehydration: Differential Thermal Analysis (DTA), X-ray and Infrared (IR). From the DTA and X-ray results, they concluded that $MgSO_3.6H_2O$ loses three water molecules between 60°C and 100°C to form $MgSO_3.3H_2O$. The latter at 200°C completely dehydrates to yield amorphus anhydrous $MgSO_3$. But in the infrared investigation, they reported that the spectra does not change when the trihydrate goes to the anhydrous state at 200°C. It does not seem plausible that the transformation suggested could have occurred without any change in the infrared spectra. The band in the 3500 cm^{-1} region is obviously an O-H stretching band, therefore, if the salt is dehydrated it would be expected that this band would disappear.

Two other investigations using DTA were carried out by groups at the Tennessee Valley Authority (TVA). These studies were not published, but are presented, in part, in an EPA-sponsored critical analysis of the Magnesia Process prepared by TVA(6). In the above cited report, Jordan's work based upon DTA leads to the conclusion that the thermal dehydration of $MgSO_3.6H_2O$ takes place in one step starting at 100°C and that $MgSO_3.3H_2O$ dehydrates in one step, starting at 160°C.

In the other study, Hatfield and co-workers reported that $MgSO_3.6H_2O$ loses nearly all its water when heated in a stream of argon or air at 104°C for 16 hrs. This, too, supports the inference that the thermal dehydration of the hexahydrate occurs in one step. It is also reported that

$MgSO_3.3H_2O$ is partially dehydrated when heated in air for 16 hrs. at 160°C.

In the above-cited EPA report, it is suggested that the apparently contradictory results may be due to differences in experimental conditions. It was speculated that the samples were heated in sealed tubes in the work of Okabe and Hori, although such information was not provided in that paper. Without presenting a critical analysis of the effects of having the samples open or closed to the atmosphere, it is concluded in the TVA report that the results of the TVA groups, in which the samples are open to the atmosphere, are valid and that $MgSO_3.6H_2O$ dehydrates in one step to $MgSO_3$ at 100°C.

The purpose of this study is to reconcile the contradictory results. Thermoanalytical techniques are used exclusively: Differential Thermal Analysis (DTA), Differential Scanning Calorimetry (DSC) and Thermal Gravimetric Analysis (TGA). Other groups in this laboratory have studied the thermal dehydration using Mass Spectroscopy and Infrared Spectroscopy. These results will be reported separately.

EXPERIMENTAL

$MgSO_3.3H_2O$ (97.9%) and $MgSO_3.6H_2O$ (99.0%) used in this study were laboratory prepared samples. Each hydrate was studied individually; also synthetic mixtures of both hydrates were studied.

The DuPont thermal analyzer was employed in this investigation. This included the DuPont 900 differential thermal analyzer (DTA) equipped with both the standard DTA cell and the differential calorimetric cell (DSC). The DuPont 950 TGA unit,which is an attachment to the DuPont 900,was also used.

DTA(9) is a thermal technique in which the heat effects, associated with chemical or physical changes, are recorded as a function of temperature or time as the substance is heated at a uniform rate. Enthalpic changes, either-endothermic or exothermic, are recorded. The sample temperature is continuously compared with a reference material temperature; the difference in temperature is recorded as a function of furnace temperature or time. DTA is reported to have been first used by LeChatelier(4) in 1887 for studying clay. Since that date many developments have been introduced and the literature has grown exponentially. DTA has been used for the study of the thermal dehydration of hydrates. For example, the reader is referred to the work of Wendlant and Hoiberg(10).

In contrast to DTA, in which the temperature difference between the sample and the reference is measured, in DSC it is the heat necessary to equalize the temperature between the sample and the reference which is measured. This technique thus can be used to measure enthalpic changes quantitatively.

TGA is a technique in which a sample is continuously weighed as it is heated at a linear rate. The resulting thermogram gives information concerning the thermal stability of the substance under investigation. TGA was first described by Honda (2) in 1915. Griffith (1) has applied TGA to the study of mixtures of hydrates and anhydrous salts.

Thermal methods of analysis are uniquely applicable for study of dehydration processes. Dehydration can be observed as endothermic changes in DTA, heats of dehydration in DSC, and as weight losses in TGA.

RESULTS AND DISCUSSION

First, the DTA results will be considered. A typical thermogram for the dehydration of $MgSO_3.3H_2O$ is shown in Figure 1. Only one endothermic transition is observed, starting at 190°C. On the other hand, in the thermogram of $MgSO_3.6H_2O$ shown in Figure 2, two endothermic transitions are observed. One starts at 90°C and the other coinciding with the endotherm observed for the trihydrate, starts at 190°C. The inference to be drawn from this data is that the hexahydrate does degrade in two steps and that the two steps involve a transition from the hexaform to the triform.

Next, studies were carried out using DSC. In this case, it is the heat input rather than the temperature which is measured. The DSC thermogram of $MgSO_3.3H_2O$ is shown in Figure 3. Only one endothermic transition was observed starting at 100°C with a peak maximum at 160°C. Both DTA and DSC analyses indicate that the thermal dehydration of $MgSO_3.3H_2O$ is a one step process. From DSC data, it appears that the dehydration of $MgSO_3.3H_2O$ starts at a low temperature, 100°C, in contrast to the DTA data from which it is inferred that 190°C is the starting temperature.

The DSC thermogram of $MgSO_3.6H_2O$ is shown in Figure 4. It is significant to note that only one endothermic transition was observed starting at 45°C with a peak maximum at 90°C. DSC results suggest that the thermal dehydration of $MgSO_3.6H_2O$

takes place in one step starting at 45°C, in contrast to the two step mechanism starting at 90°C to be drawn from the DTA results.

Since the phenomena measured by DTA and DSC is the same, it was expected that the results would be consistent. Surprisingly, this was not the case. From the DSC data, it appears that the dehydration of both hydrates starts at a lower temperature and the hexaform dehydrates in one step and not in two steps as was inferred from the DTA results.

The apparent contradiction between the DTA and the DSC results can be rationalized by considering the relationship between the dehydration reaction and the sample environmental conditions. In any dehydration reaction water is liberated; if the latter is continuously removed from the reaction atmosphere two consequences are observed. First, the dehydration starts at a lower temperature, and, secondly, equilibrium is not attained.

On the one hand in the DSC studies, the sample is placed in an open dish and is heated under a sweeping stream of nitrogen. Under these conditions the liberated water is continuously removed from the reaction environment. On the other hand, in the DTA studies, the sample is heated in a self-generated atmosphere because it is placed in a capillary tube. Under the open conditions encountered in DSC, thermal dehydration starts at a lower temperature, 45°C versus 90°C and only one endothermic transition is observed for the hexaform, whereas in the DTA studies two were observed.

To confirm the above rationalization of the differences between the DTA and the DSC studies, the thermal dehydration was studied by another independent technique, TGA, in which it was possible to heat the samples either in an open condition or in a self-generated atmosphere. TGA thermograms, under open conditions, were obtained by the conventional procedure in which the sample is placed in an open platinum dish under a sweeping blanket of nitrogen. TGA, under self-generated atmosphere, was achieved by placing the sample in a capillary tube with a thermocouple inside. Then, the whole tube was placed in the platinum dish.

In Figure 5, the TGA of $MgSO_3.3H_2O$, under open conditions is shown. Thermal dehydration starts at 100°C and takes place in one step. The weight loss of the sample is 34.0%, which corresponds to the loss of 3 moles of water. The effect of operating under open conditions is shown in Figure 6 for $MgSO_3.6H_2O$. Thermal dehydration of the hexaform starts at 70°C and takes place in one step. The weight loss of the sample is 51%, which corresponds to the loss of 6 moles of water.

The thermograms for the trihydrate and the hexahydrate, respectively, under the conditions of a self-generated atmosphere are shown in Figures 7 and 8. $MgSO_3.3H_2O$, as shown in Figure 7, loses 34.5% of its weight in one step. This is in a very good agreement with TGA under open conditions, but thermal dehydration starts at 220°C, i.e., at a higher temperature, because of the self-generated atmosphere conditions. The weight loss for the hexahydrate under a self-generated atmosphere, as shown in Figure 8, is 51%, which corresponds to the loss of 6 moles of water. It is significant to note that the thermal dehydration of the hexaform starts at a higher temperature and takes place in two steps, with a weight loss of 25.5% in each step. In other words, $MgSO_3.6H_2O$ loses 3 moles of water in each dehydration step.

Thus, when the TGA study of the hexaform is carried out under a non-equilibrium condition, because of heating the sample under a sweeping blanket of nitrogen, equilibrium is not attained due to the continuous removal of the liberated water. As a result, thermal dehydration starts at a lower temperature and takes place in one step. On the other hand, if equilibrium is attained, because of heating the sample in a self-generated atmosphere, the TGA results showed that thermal dehydration starts at a higher temperature and takes place in two steps. First, dehydration leads to the formation of $MgSO_3.3H_2O$, then, to the anhydrous $MgSO_3$. These TGA observations are consistent with the explanation of the differences observed in the DTA and the DSC studies. The results are summarized in Table 1 for $MgSO_3.3H_2O$ and in

Table 2 for $MgSO_3.6H_2O$. The onset of the endotherms as a function of the method are compared.

APPLICATION

The observations made in the TGA studies were used as bases for developing a new analytical method for quantitating mixtures of the tri- and hexahydrates. It had been observed that although the second weight loss for the hexahydrate overlapped that of the trihydrate, nevertheless, from a knowledge of the first weight loss of the hexahydrate, the contribution of the hexahydrate to the second weight loss could be calculated.

The accuracy of the method was investigated by analyzing the thermograms of synthetic mixtures of the hexahydrate and the trihydrate ranging from 10% to 90%. The thermograms are shown in Figures 9-19.

From these thermograms, the water content and the magnesium sulfite content could be calculated. First, the water content calculations will be considered and compared to the theoretical values. The water content of each hydrate in a mixture can be obtained from the TGA thermogram, by taking into account the fact that the weight loss in the first dehydration step at 175°C represents the first three moles of water in $MgSO_3.6H_2O$, i.e., 50% of the water content of the hexaform. Thus the water content can be calculated, from the TGA thermogram, as follows:

$$\%\ H_2O \text{ in } MgSO_3.6H_2O \text{ in a mixture} = (\%\text{ weight loss in the first step at } 175^\circ C)(2)$$

% H_2O in $MgSO_3.3H_2O$ in a mixture = (% weight loss in the second step at 400° C) - (% weight loss in the first step at 175° C)

The theoretical values of the water content are calculated as follows:

% H_2O in $MgSO_3.6H_2O$ in a mixture = % $MgSO_3.6H_2O$ in the mixture $\times \frac{6H_2O}{MgSO_3.6H_2O} \times 100$

= % $MgSO_3.6H_2O$ in the mixture $\times \frac{108}{212.3} \times 100$

= % $MgSO_3.6H_2O$ in the mixture $\times 50.9$

% H_2O in $MgSO_3.3H_2O$ in a mixture = % $MgSO_3.3H_2O$ in the mixture $\times \frac{3\ H_2O}{MgSO_3.3H_2O} \times 100$

= % $MgSO_3.3H_2O$ in the mixture $\times \frac{54}{158.3} \times 100$

= % $MgSO_3.3H_2O$ in the mixture $\times 34.1$

Both theoretical and observed values of the water content are shown in Table 3. The percent water (observed values) is plotted versus magnesium sulfite content in Figure 21 for $MgSO_3.3H_2O$ and in Figure 22 for $MgSO_3.6H_2O$.

The magnesium sulfite content can be calculated from the TGA thermograms by considering the same aspects mentioned in the water content calculations, and, taking into account that 6 moles of water represent 50.9% by weight of $MgSO_3.6H_2O$, and that 3 moles of water represent 34.1% by weight of $MgSO_3.3H_2O$. Thus the percent of each hydrate in a mixture is calculated as follows:

$$\%\ MgSO_3.6H_2O \text{ in a mixture} = \%\text{ weight loss at } 175^\circ C \times 2 \times \frac{100}{50.9}$$

$$\%\ MgSO_3.3H_2O \text{ in a mixture} = (\%\text{ weight loss between } 175^\circ \text{ and } 400^\circ C - \%\text{ weight loss at } 175^\circ C) \times \frac{100}{34.1}$$

The data obtained is compared versus the theoretical values in Table 4 for the triform and in Table 5 for the hexaform. The accuracy of the method, as shown in Tables 4 and 5, was found to be ± 3% if no calibration curve is used. A value of less than 1% is to be expected if a calibration curve is used.

Furthermore, the TGA method was tested in the presence of TGA inactive material. This was achieved by analyzing a mixture of $MgSO_3.3H_2O$, $MgSO_3.6H_2O$ and glass beads. The mixture was subjected to the same experimental conditions of a self-generated atmosphere. The TGA thermogram is shown in Figure 20. From the results summarized in Table 6, it is concluded that this method does not suffer any interferences due to the presence of thermally inactive materials.

Relevant to the nature of the materials obtained in the Magnesia Process, the quantitative analysis of mixtures of $MgSO_3.3H_2O$ and $MgSO_3.6H_2O$ has to be considered in the light of the following contexts: one, only these two hydrates are present; two, in the presence of MgO, three, in the presence of another hydrate, $MgSO_4.7H_2O$. The latter is a common product in the Magnesia Process.

When only the two hydrates are present, methods available include total sulfite by iodine titration(3), X-ray analysis(5), and a wet chemical method developed by Dr. Ray which will be referred to as Ray's Method(7).

In the presence of MgO and/or $MgSO_4.7H_2O$ the measurement of the sulfite content cannot be used to determine the ratio of the hydrates. From the available literature which is limited to the one paper previously cited(5), it is not clear as to whether or not quantitative analysis is possible using the X-ray method in the presence of MgO and/or $MgSO_4.7H_2O$. Ray's method is applicable in the presence of these materials.

The TGA method is simpler than Ray's method but it is not applicable in the presence of $MgSO_4.7H_2O$. On the other hand, Dr. Ray(7) has developed a simple procedure for quantitatively stripping $MgSO_4.7H_2O$ from mixtures of $MgSO_3.3H_2O$, $MgSO_3.6H_2O$ and MgO. Therefore it appears that the TGA technique in conjunction with Ray's stripping procedure is the optimum method.

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Table 1

Starting Temperature of Thermal Dehydration of $MgSO_3.3H_2O$

Technique	Experimental Conditions	Onset of the Endotherm
DTA	self-generated	190°C
DSC	open conditions	100°C
TGA	open conditions	100°C
TGA	self-generated	200°C

Table 2

Starting Temperature of Thermal Dehydration of $MgSO_3 \cdot 6H_2O$

Technique	Experimental Condition	First Endotherm	Second Endotherm
DTA	self-generated	90°C	190°C
DSC	open conditions	45°C	not observed
TGA	open conditions	70°C	not observed
TGA	self-generated	125°C	220°C

Table 3

Calculated and Observed Values

of Water Content in

Synthetic Mixtures

Std#	Composition		Calculated H_2O content in			Observed H_2O content in		
	%$MgSO_3.3H_2O$	%$MgSO_36H_2O$	Triform	Hexaform	Total	Triform	Hexaform	Total
1	0.0%	100%	0%	50.9%	50.9%	0%	52%	52%
2	10.29%	84.71%	3.5%	45.56%	49.06%	3.4%	44.4%	47.8%
3	18.96%	81.04%	6.46%	41.19%	47.65%	6.6%	39.4%	46%
4	29.37%	70.63%	10%	35.9%	45.9%	9.0%	35.0%	44%
5	38.24%	61.76%	13.03%	31.39%	44.42%	12%	32%	44%
6	50%	50%	17.04%	25.42%	42.46%	16.4%	26%	42.4%
7	59.84%	40.16%	20.39%	20.41%	40.8%	19.8%	20.4%	40.2%
8	70.19%	29.81%	23.91%	15.15%	39.06%	23%	16%	39%
9	79.34%	20.66%	27.03%	10.5%	37.53%	26.4%	10.6%	37%
10	90.18	9.82%	30.72%	4.99%	35.71%	30.2%	5%	35.2%
11	100%	0%	34.1%	0%	34.1%	34.8%	0%	34.8%

Table 4

$MgSO_3.3H_2O$ Content from TGA

Std#	Theoretical Value	Observed Value	Deviation
1	0%	0%	0%
2	10.29%	10.26%	-.03%
3	18.96%	19.35%	+.39%
4	29.37%	26.39%	-2.98%
5	38.24%	35.24%	-3.00%
6	50.0%	48.1%	-1.9%
7	59.84%	58.06%	-1.78%
8	70.19%	67.45%	-2.74%
9	79.34%	77.42%	-1.92%
10	90.18%	88.56%	-1.62%
11	100.00%	102.0%	+2.0%

Table 5

$MgSO_3.6H_2O$ Content

From TGA

Std#	Theoretical Value	Observed Value	Deviation
1	100%	102.2%	+2.2%
2	89.71%	87.3%	-2.41%
3	81.04%	78.45%	-2.54%
4	70.63%	68.8%	-1.83%
5	61.76%	62.9%	+1.14%
6	50%	51.11%	+1.11%
7	40.16%	40.1%	-.06%
8	29.81%	31.45%	+1.64%
9	20.66%	20.84%	+.18%
10	9.82%	9.83%	+.01%
11	0%	0%	0%

Table 6

TGA results of a synthetic mixture containing thermally inactive compound

	Theoretical Value	Observed Value
$MgSO_3.3H_2O$	44.3%	43.2%
$MgSO_3.6H_2O$	30.0%	29.3%
glass beads	25.7%	27.5%*

*This value was obtained by difference

DUPONT INSTRUMENTS

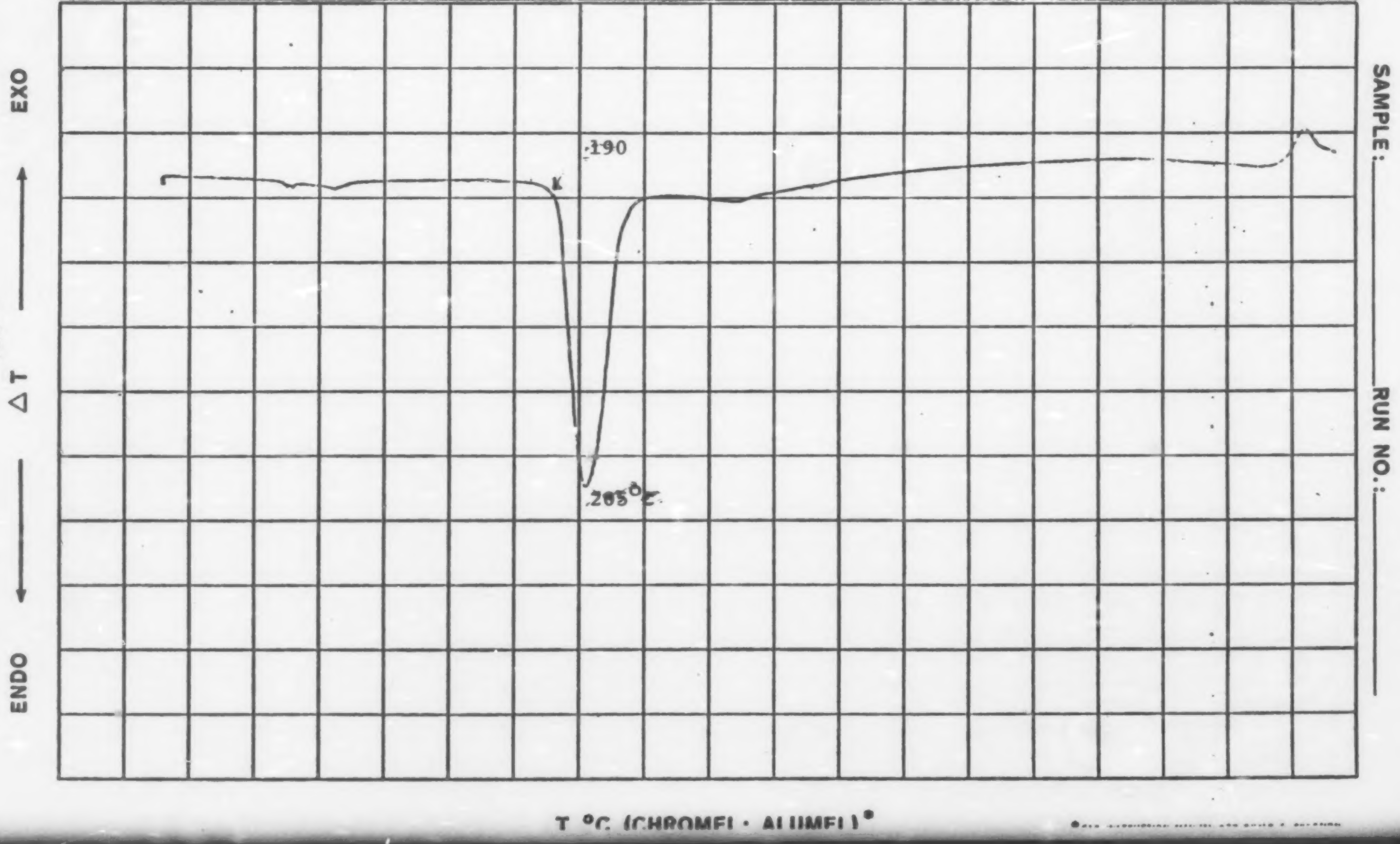
SAMPLE: $MgSO_3 . 3H_2O$
Fig. No. 1
ORIGIN:

SIZE 6 mg
REF. glass beads
PROG. MODE heat
RATE 10 °C/MIN. ,START 30 °C

ATM. self-generated MM

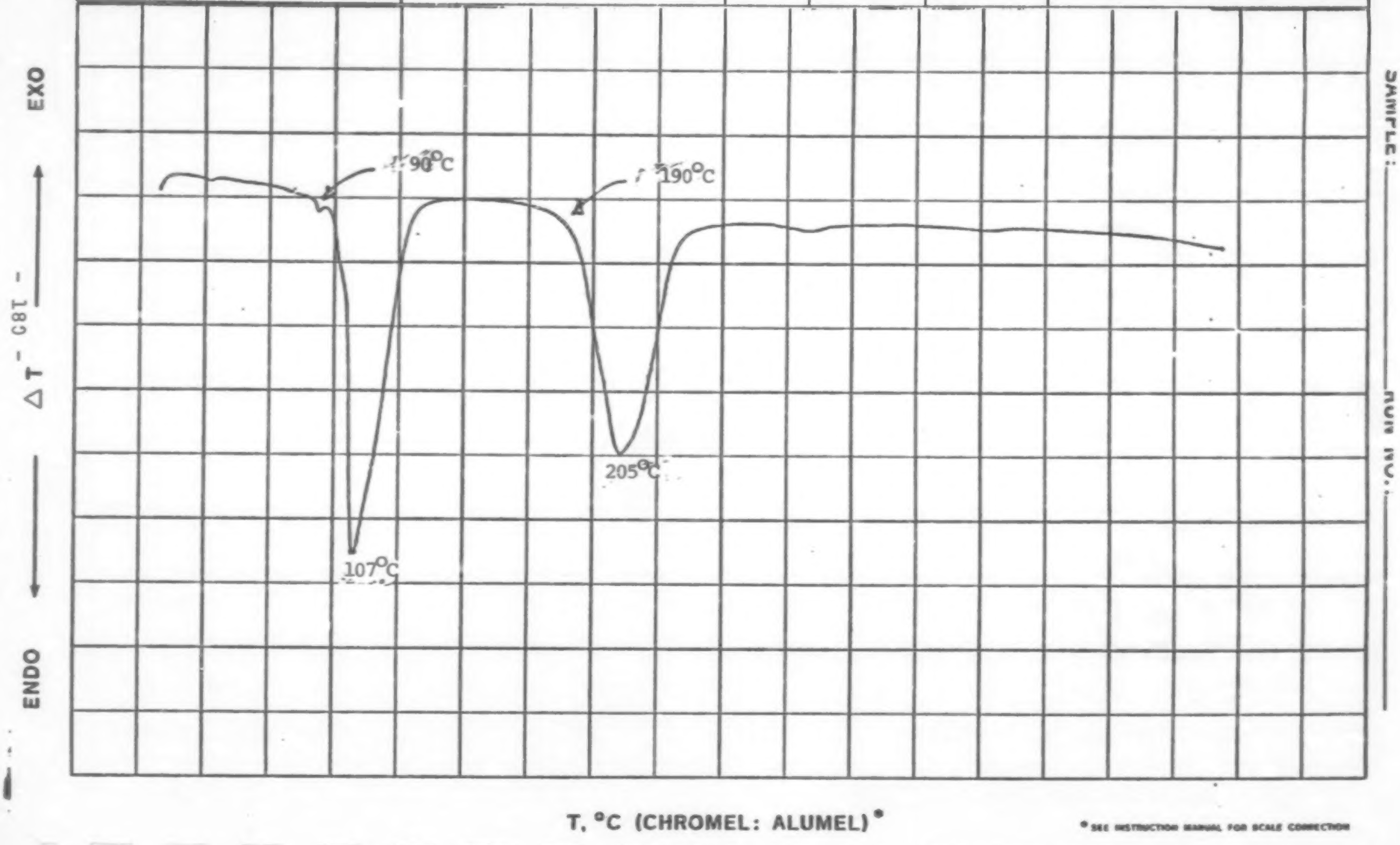
	T	ΔT
SCALE	50 °C/IN.	2 °C/IN.
SHIFT	0 IN.	0 IN.

RUN NO. 1
DATE 11/8/74
OPERATOR RM
BASE LINE SLOPE 0



DU PONT INSTRUMENTS

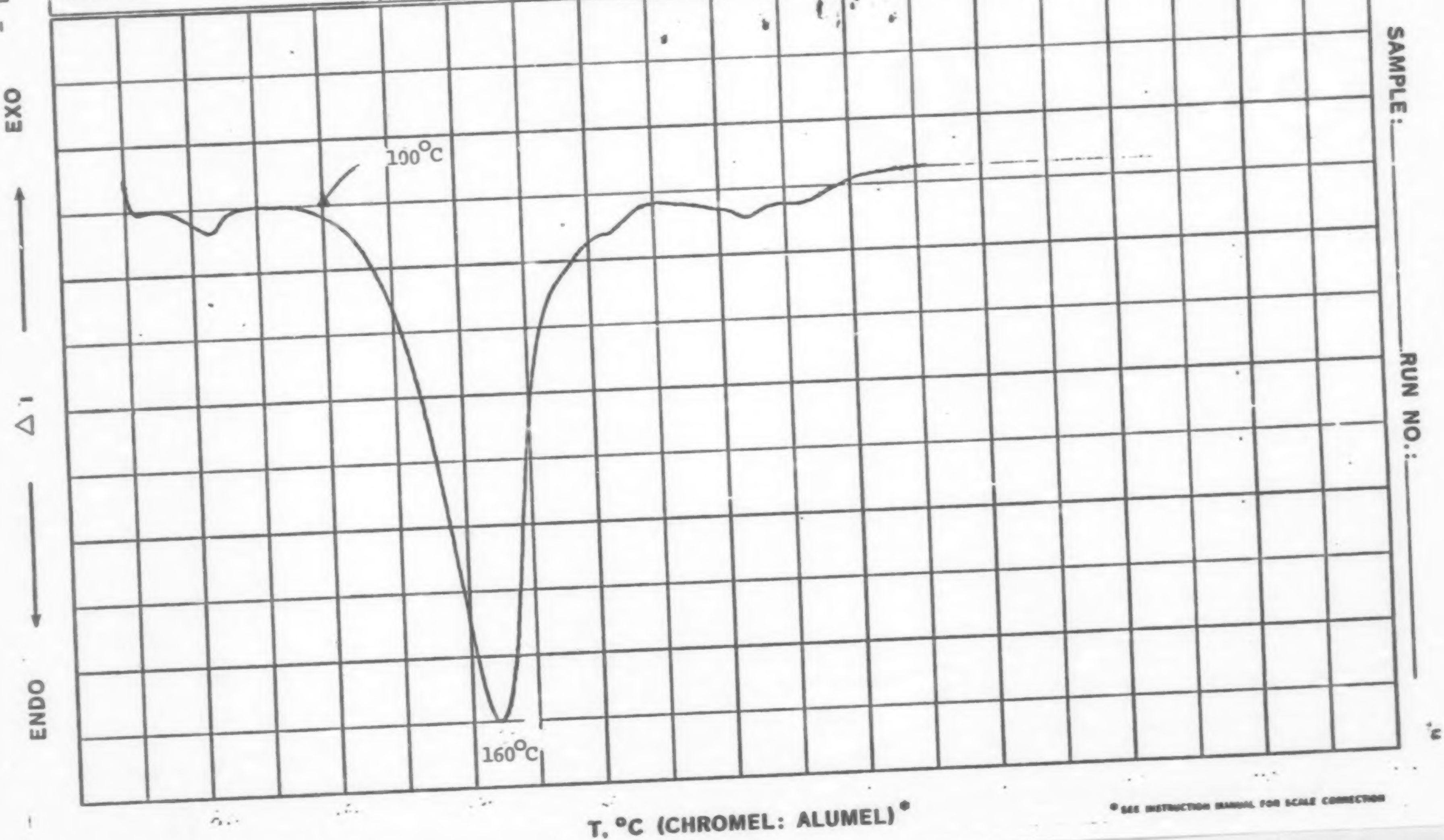
SAMPLE: $MgSO_3 . 6H_2O$	SIZE 8 mg	ATM. self-generated MM	RUN NO. 1
Fig. No. 2	REF. glass beads	T / ΔT	DATE 11/8/74
	PROG. MODE heat	SCALE 50 °C/IN. / 2 °C/IN.	OPERATOR RM
ORIGIN:	RATE 10 °C/MIN, START 30 °C	SHIFT 0 IN. / 0 IN.	BASE LINE SLOPE 0



* SEE INSTRUCTION MANUAL FOR SCALE CORRECTION

DUPONT INSTRUMENTS

SAMPLE: $MgSO_3 . 3H_2O$ Fig. No. 3	SIZE 12 mg	ATM. N_2 30 MM	RUN NO. 1
	REF. empty pan	T / ΔT	DATE 11/9/79
	PROG. MODE heat	SCALE 50 °/in. / 1 °/in.	OPERATOR RM
ORIGIN:	RATE 10 °/min, START 25 °C	SHIFT 0 IN. / 0 IN.	BASE LINE SLOPE 0



SAMPLE: ____________ RUN NO.: ____________

* SEE INSTRUCTION MANUAL FOR SCALE CORRECTION

DUPONT INSTRUMENTS

SAMPLE: $MgSO_3 \cdot 6H_2O$

Fig. No. 4

ORIGIN:

SIZE 5 mg

REF. empty pan

PROG. MODE heat

RATE 10 °C/min, START 30 °C

ATM. N_2, 30 MM

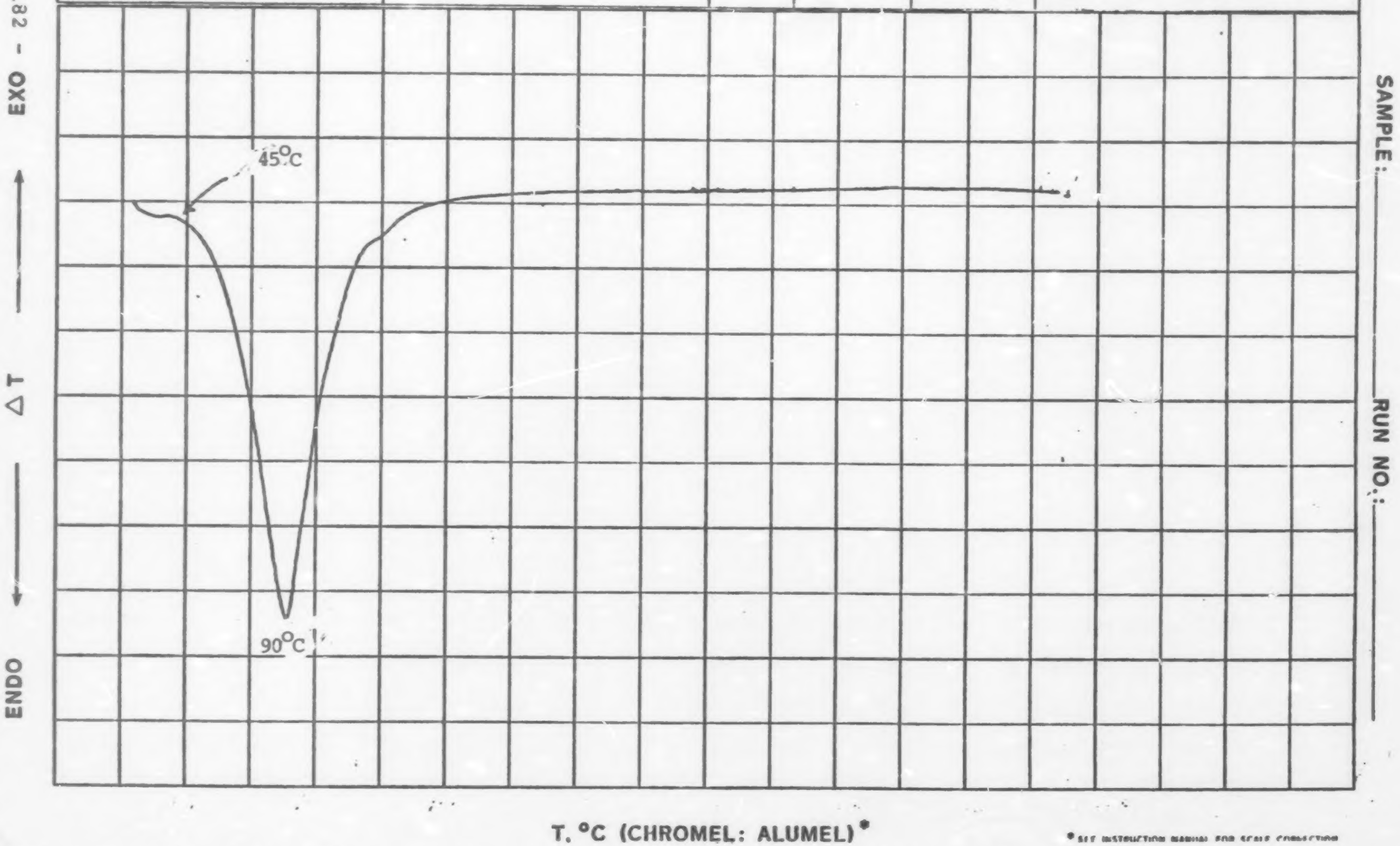
	T	ΔT
SCALE	50 °C/in	1 °C/in
SHIFT	0 IN.	0 IN.

RUN NO. 1

DATE 11/9/74

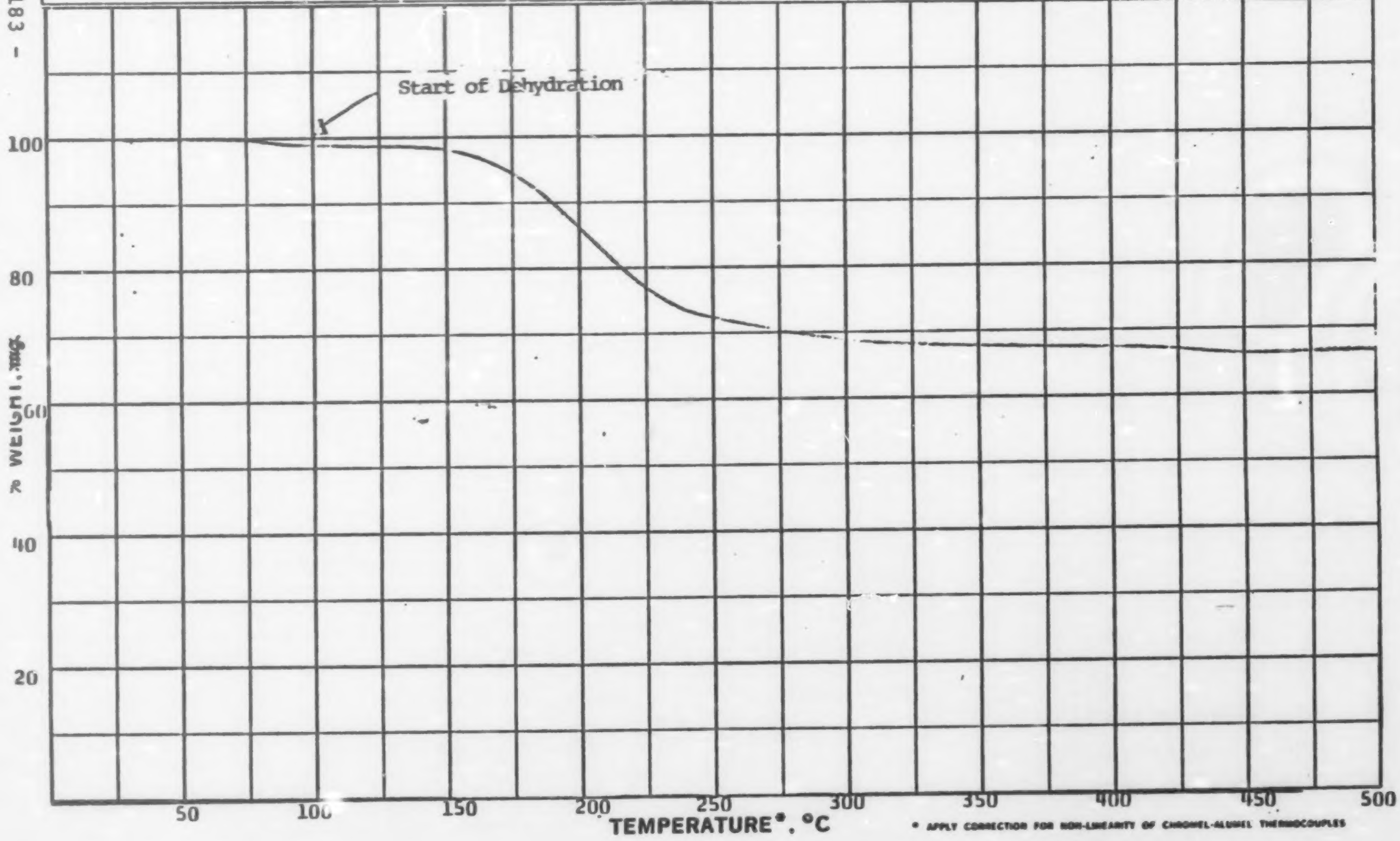
OPERATOR RM

BASE LINE SLOPE 0



DUPONT INSTRUMENTS

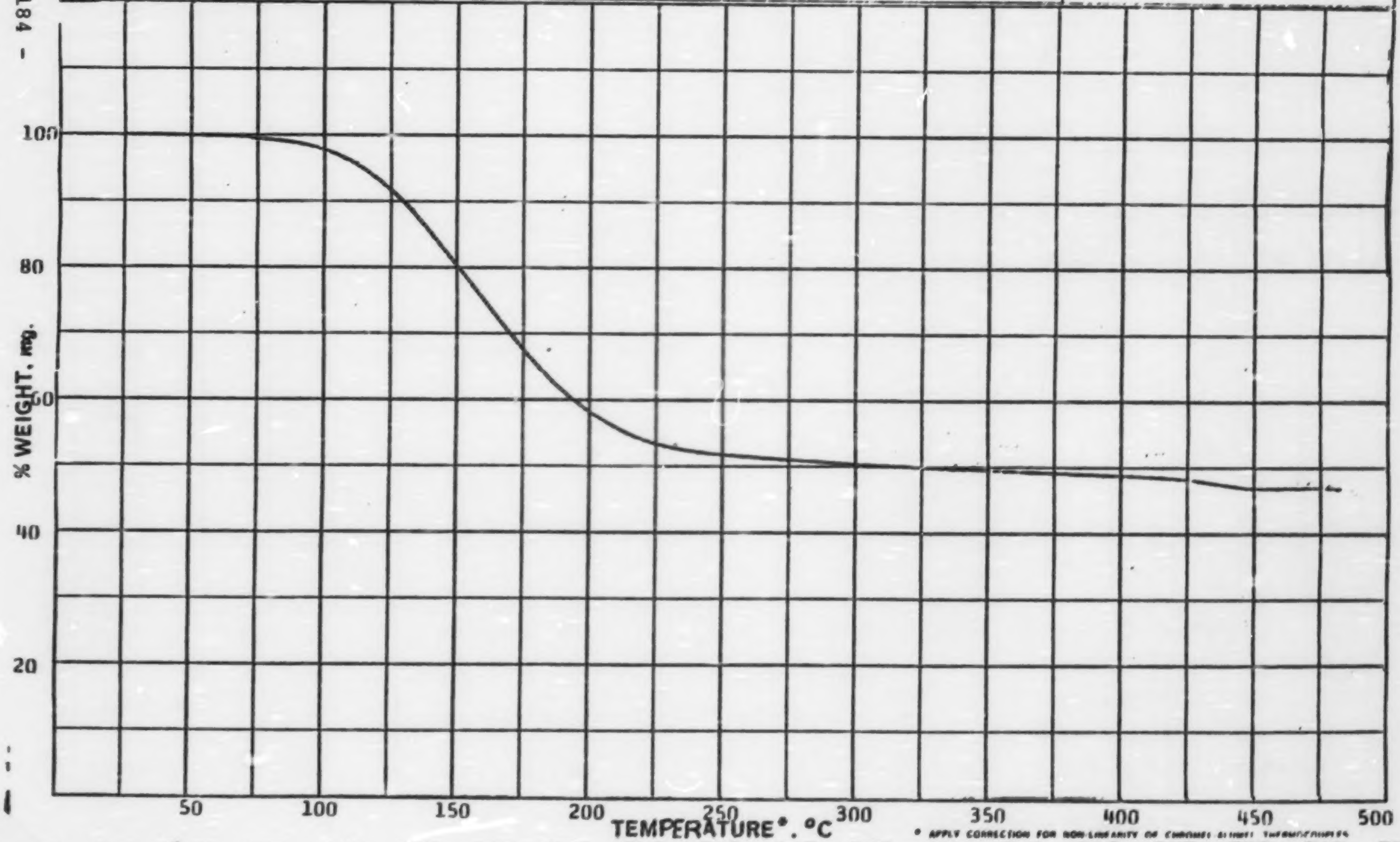
SAMPLE: $MgSO_3 . 3H_2O$ Fig. No. 5 SIZE 20 mg.	X-AXIS TEMP. SCALE 50 °C/inch SHIFT 0 inch TIME SCALE (ALT.)	Y-AXIS SCALE 4 mg./inch (SCALE SETTING X 2) SUPPRESSION 50 mg.	RUN NO. 1 DATE 11/12/74 OPERATOR RM HEATING RATE 10 °C/min. ATM. N_2, 30 ml/min. TIME CONSTANT 1 sec.



* APPLY CORRECTION FOR NON-LINEARITY OF CHROMEL-ALUMEL THERMOCOUPLES

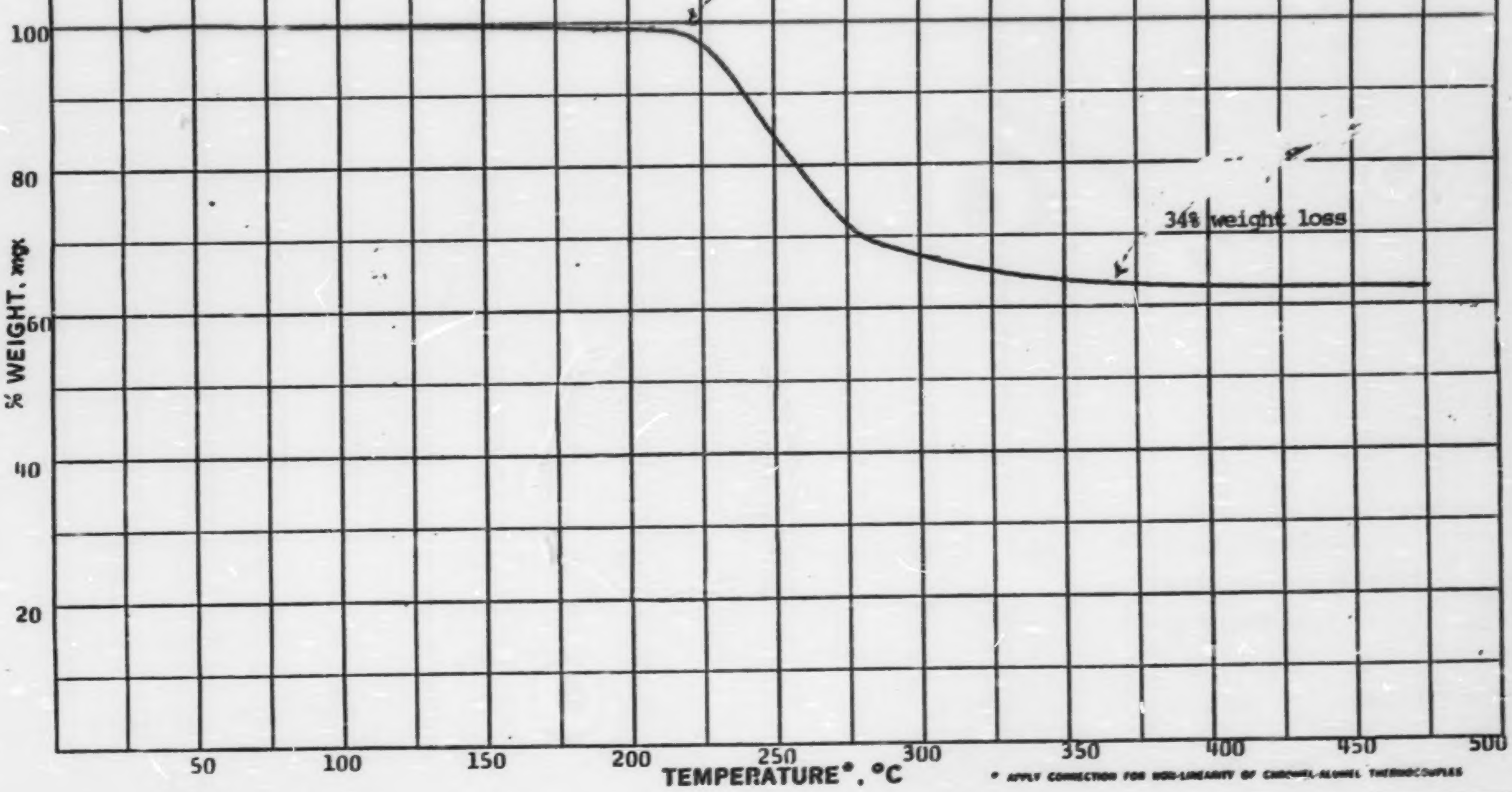
DUPONT INSTRUMENTS

SAMPLE: $MgSO_3.6H_2O$	**X-AXIS**	**Y-AXIS**	**RUN NO.** 1 **DATE** 11/12/7[illegible]
Fig. No. 6	**TEMP. SCALE** 50 °C/inch	**SCALE** 4 mg./inch	**OPERATOR** RM
	SHIFT 0 inch	(SCALE SETTING X 2)	**HEATING RATE** 10 °C/min.
			ATM. N_2, 30 ml/min.
SIZE 20 mg.	**TIME SCALE (ALT.)**	**SUPPRESSION** 5 mg.	**TIME CONSTANT** 1 sec.



INSTRUMENTS

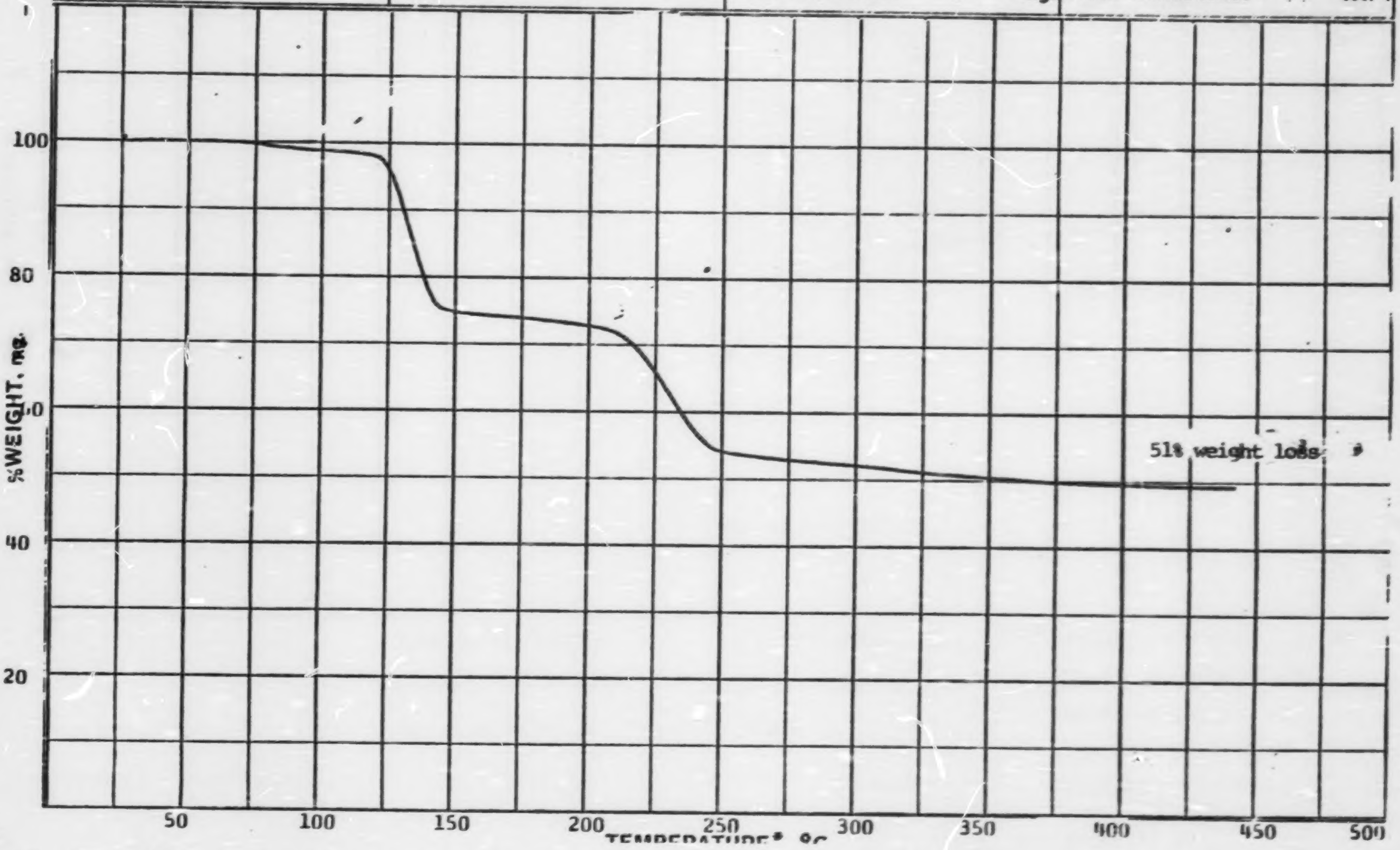
SAMPLE: $MgSO_3 \cdot 3H_2O$	X-AXIS	Y-AXIS	RUN NO. 1 DATE 11/15/79
Fig. No. 7	TEMP. SCALE 50 °C/inch	SCALE 2 mg./inch (SCALE SETTING X 2)	OPERATOR RM
	SHIFT 0 inch		HEATING RATE 10 °C/min.
			ATM. self-generated
SIZE 5 mg.	TIME SCALE (ALT.)	SUPPRESSION 60 mg.	TIME CONSTANT 1 sec.



* APPLY CORRECTION FOR NON-LINEARITY OF CHROMEL-ALUMEL THERMOCOUPLES

DU PONT INSTRUMENTS

SAMPLE: $MgSO_3 \cdot 6H_2O$	**X-AXIS**	**Y-AXIS**	**RUN NO.** 1 **DATE** 11/16/7
Fig. No. 8	**TEMP. SCALE** 50 °C/inch	**SCALE** 1 mg./inch (SCALE SETTING X 2)	**OPERATOR** RM
	SHIFT 0 inch		**HEATING RATE** 10 °C/min.
			ATM. self-generated
SIZE 5 mg.	**TIME SCALE (ALT.)**	**SUPPRESSION** 60 mg.	**TIME CONSTANT** sec.



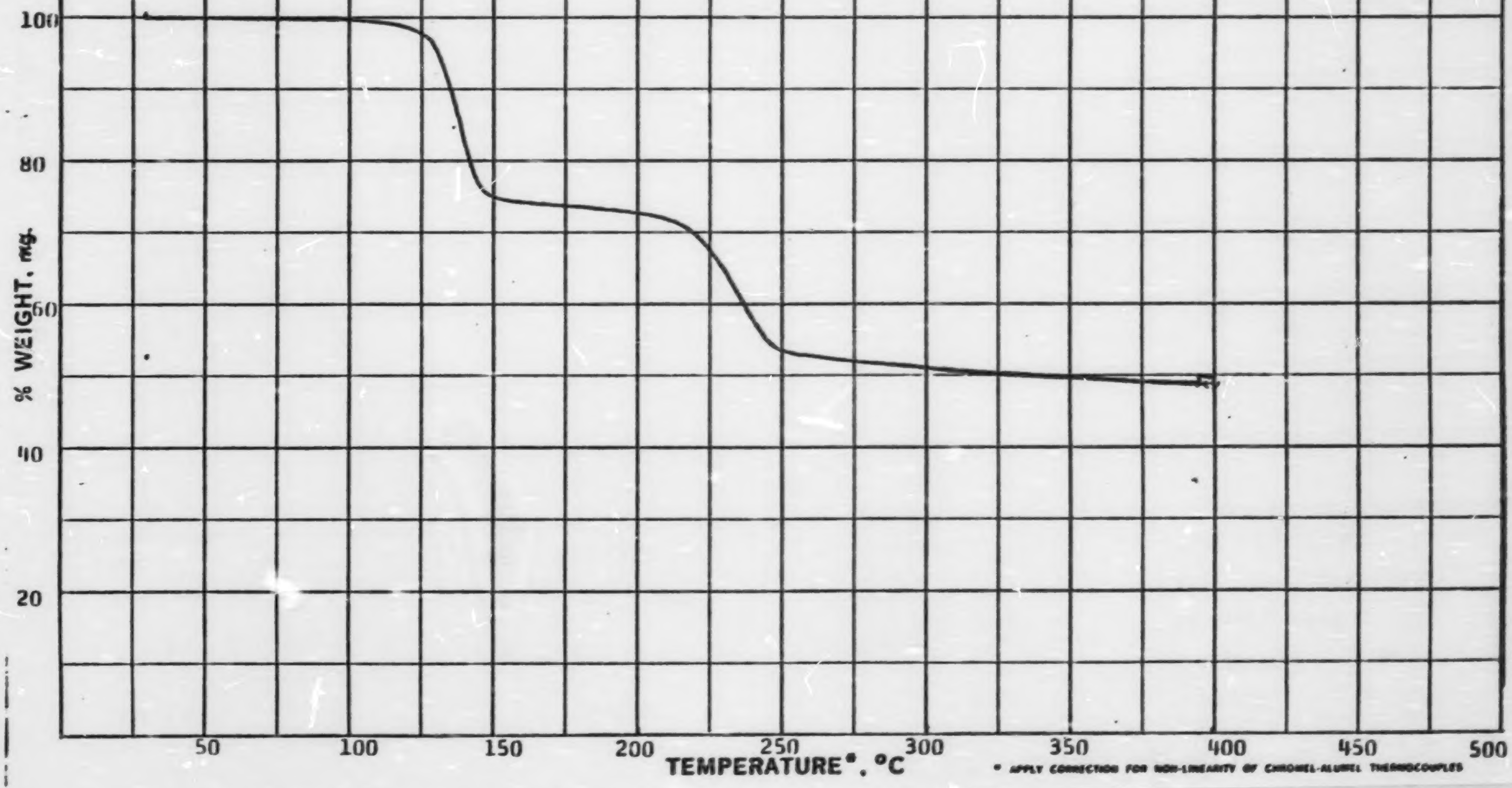
DU PONT INSTRUMENTS

SAMPLE: Std #1
100% $MgSO_3 \cdot 6H_2O$
Fig. No. 9
SIZE 12 mg.

X-AXIS
TEMP. SCALE 50 °C/inch
SHIFT 0 inch
TIME SCALE (ALT.)

Y-AXIS
SCALE 2 mg./inch
(SCALE SETTING X 2)
SUPPRESSION 50 mg.

RUN NO. 1 **DATE** 11/17/7
OPERATOR RM
HEATING RATE 10 °C/min.
ATM. self-generated
TIME CONSTANT 1 sec.



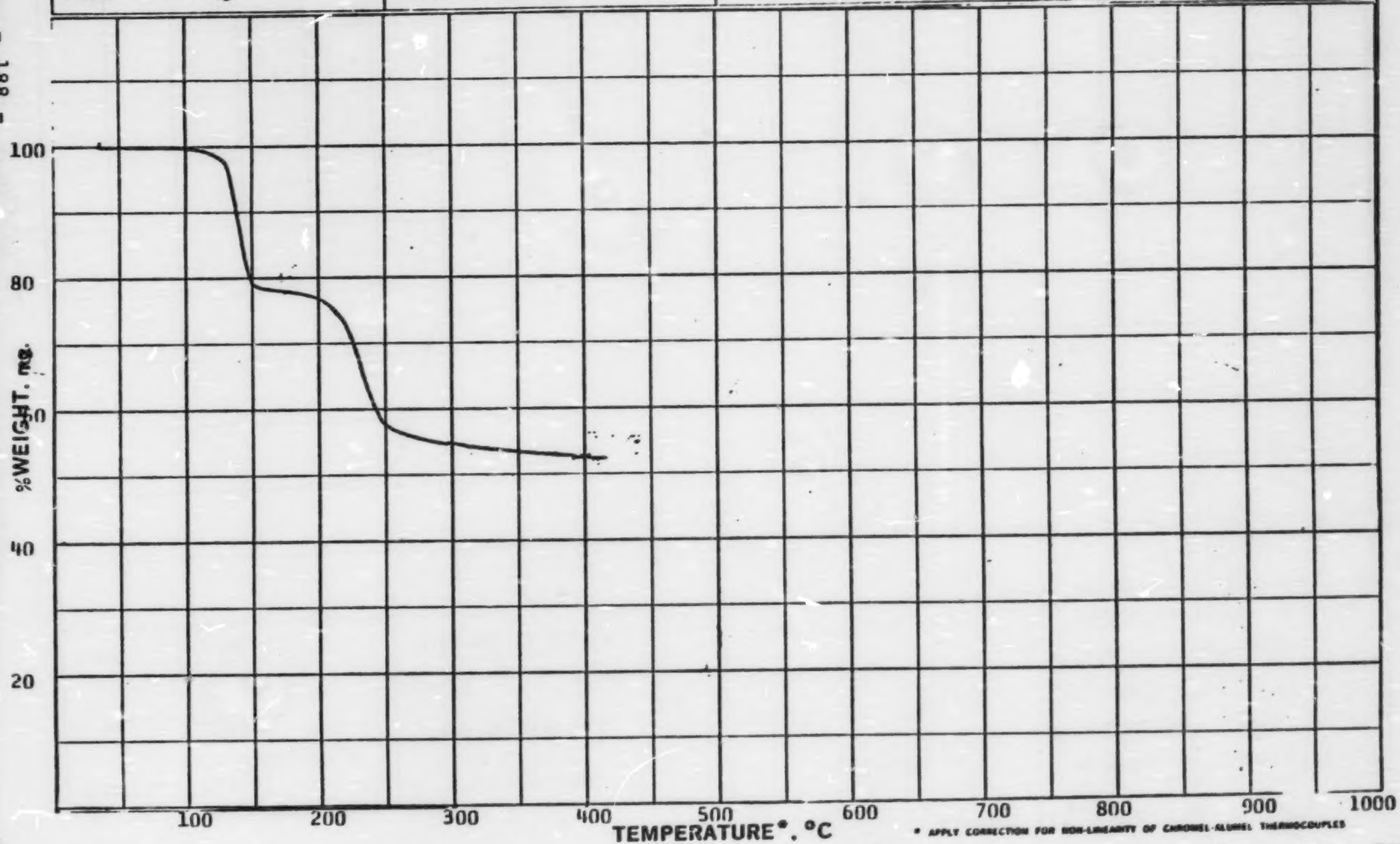
DU PONT INSTRUMENTS

SAMPLE: Std #2
10.29% $MgSO_3.3H_2O$
89.71% $MgSO_3.6H_2O$
Fig. No. 10
SIZE 10 mg.

X-AXIS
TEMP. SCALE 100 °C/inch
SHIFT 0 inch
TIME SCALE (ALT.)

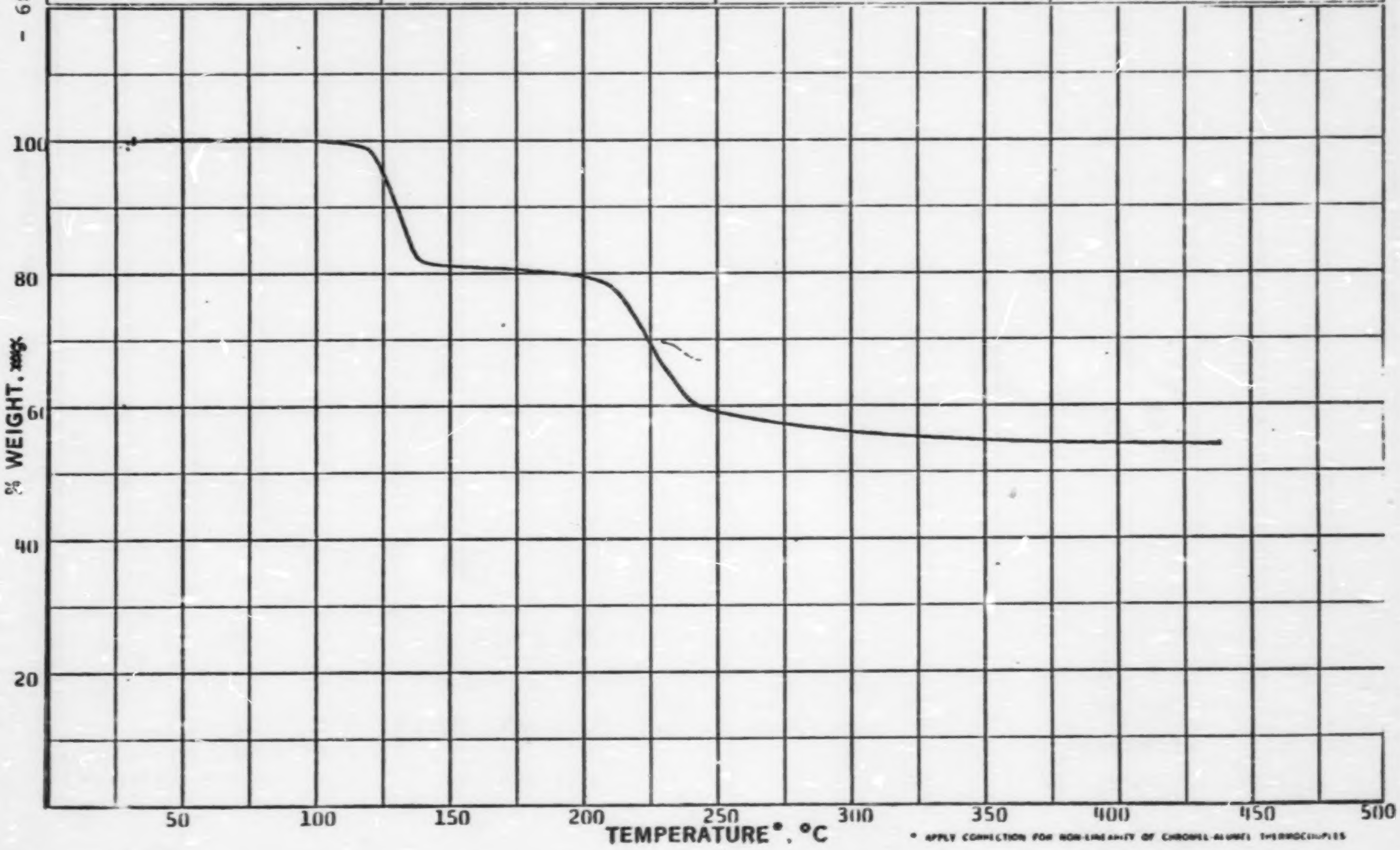
Y-AXIS
SCALE 2 mg./inch
(SCALE SETTING X 2)
SUPPRESSION 50 mg.

RUN NO. 1 **DATE** 11/20/74
OPERATOR RM
HEATING RATE 10 °C/min.
ATM. self-generated
TIME CONSTANT 1 sec.



DU PONT INSTRUMENTS

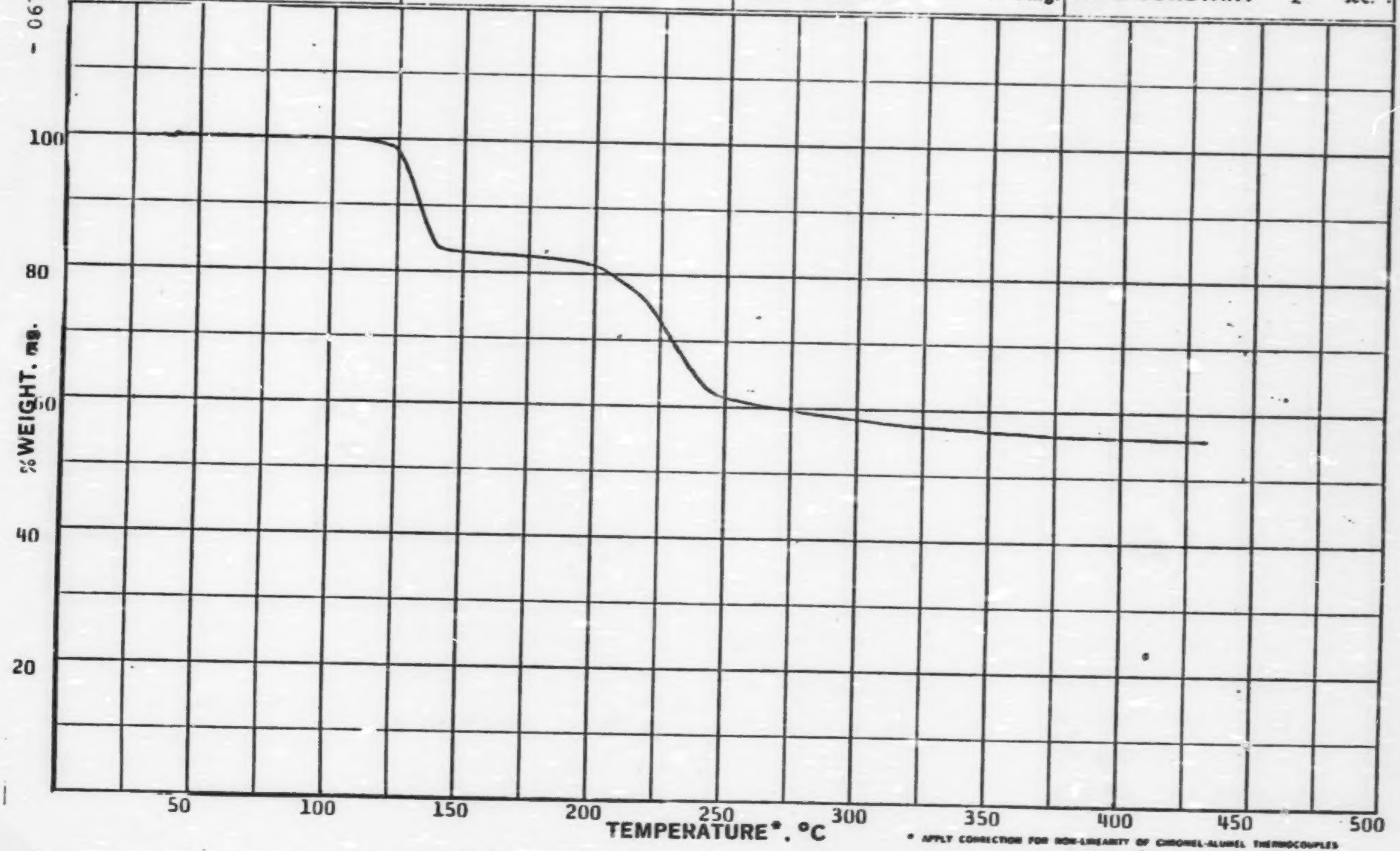
SAMPLE: Std #3 18.96% $MgSO_3 \cdot 3H_2O$ 81.04% $MgSO_3 \cdot 6H_2O$ Fig. No. 11 **SIZE** 15 mg.	**X-AXIS** **TEMP. SCALE** 50 °C/inch **SHIFT** 0 inch **TIME SCALE (ALT.)**	**Y-AXIS** **SCALE** 2 mg./inch (SCALE SETTING X 2) **SUPPRESSION** 50 mg.	**RUN NO.** 1 **DATE** 11/21/7 **OPERATOR** RM **HEATING RATE** 10 °C/min. **ATM.** self-generated **TIME CONSTANT** 1 sec.



* APPLY CORRECTION FOR NON-LINEARITY OF CHROMEL-ALUMEL THERMOCOUPLES

DUPONT INSTRUMENTS

SAMPLE: Std #4 29.37% $MgSO_3 \cdot 3H_2O$ 70.63% $MgSO_3 \cdot 6H_2O$ Fig. No. 12 **SIZE** 10 mg.	**X-AXIS** **TEMP. SCALE** 50 °C/inch **SHIFT** 0 inch **TIME SCALE (ALT.)**	**Y-AXIS** **SCALE** 2 mg./inch (SCALE SETTING X 2) **SUPPRESSION** 50 mg.	**RUN NO.** 1 **DATE** 11/22/7 **OPERATOR** RM **HEATING RATE** 10 °C/min. **ATM.** self-generated **TIME CONSTANT** 1 sec.



* APPLY CORRECTION FOR NON-LINEARITY OF CHROMEL-ALUMEL THERMOCOUPLES

DU PONT INSTRUMENTS

SAMPLE. Std #5
38.24% $MgSO_3 \cdot 3H_2O$
61.76% $MgSO_3 \cdot 6H_2O$

Fig. No. 13

SIZE 12 mg.

X-AXIS

TEMP. SCALE 50 °C/inch

SHIFT 0 inch

TIME SCALE (ALT.)

Y-AXIS

SCALE 2 mg./inch

(SCALE SETTING X 2)

SUPPRESSION 50 mg.

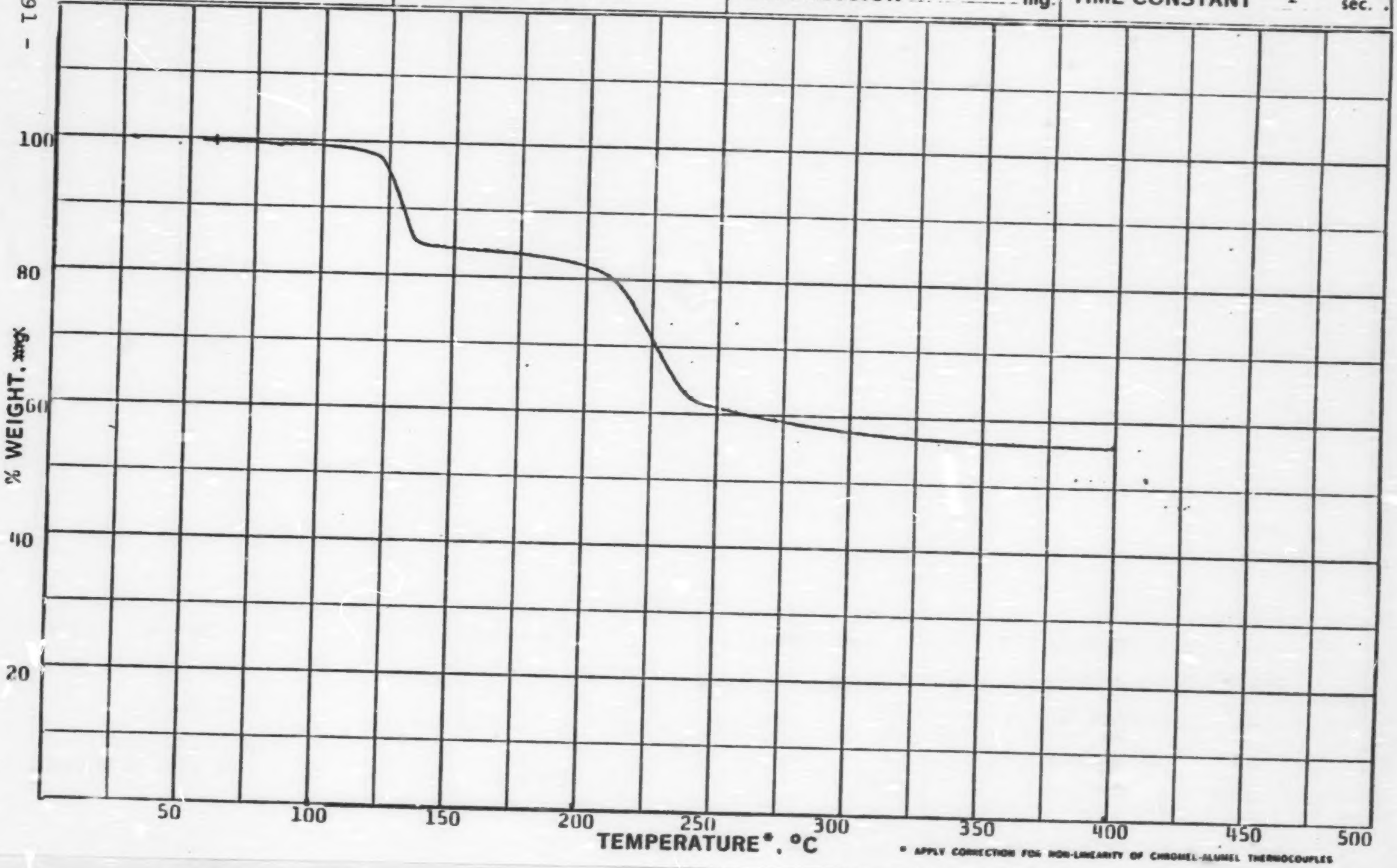
RUN NO. 1 DATE 11/25/74

OPERATOR RM

HEATING RATE 10 °C/min.

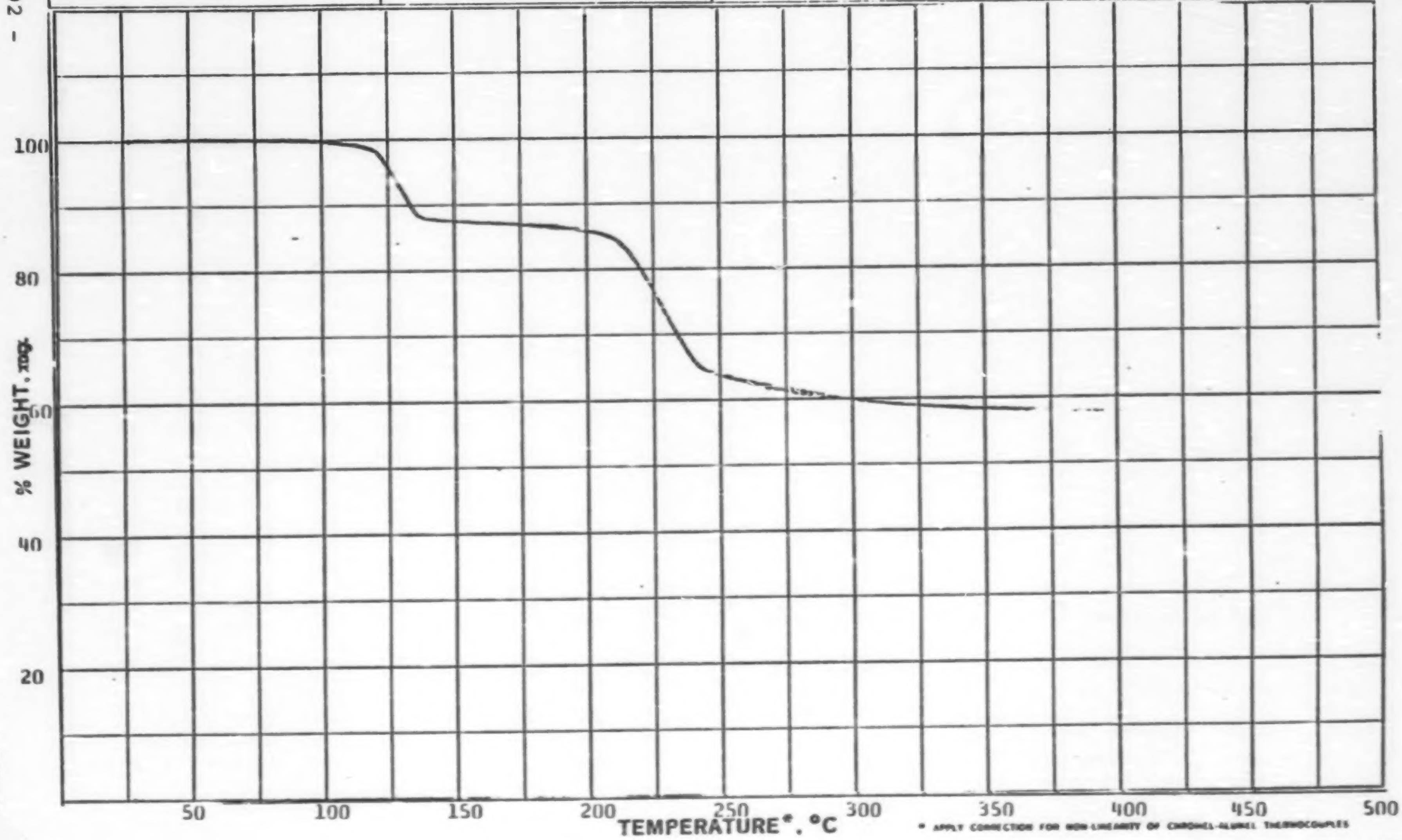
ATM. self-generated

TIME CONSTANT 1 sec.



DU PONT INSTRUMENTS

SAMPLE: Std #6 50% $MgSO_3 \cdot 3H_2O$ 50% $MgSO_3 \cdot 6H_2O$	**X-AXIS**	**Y-AXIS**	**RUN NO.** 1 **DATE** 11/25/74
	TEMP. SCALE 50 °C/inch	**SCALE** 2 mg./inch (SCALE SETTING X 2)	**OPERATOR** RM
Fig. No. 14	**SHIFT** 0 inch		**HEATING RATE** 10 °C/min.
			ATM. self-generated
SIZE 10 mg.	**TIME SCALE (ALT.)**	**SUPPRESSION** mg.	**TIME CONSTANT** 1 sec.



* APPLY CORRECTION FOR NON-LINEARITY OF CHROMEL-ALUMEL THERMOCOUPLES

DU PONT INSTRUMENTS

SAMPLE: Std #7
59.84% $MgSO_3 \cdot 3H_2O$
40.16% $MgSO_3 \cdot 6H_2O$

Fig. No. 15

SIZE 12 mg.

X-AXIS

TEMP. SCALE 50 °C/inch

SHIFT 0 inch

TIME SCALE (ALT.)

Y-AXIS

SCALE 2 mg./inch

(SCALE SETTING X 2)

SUPPRESSION 50 mg.

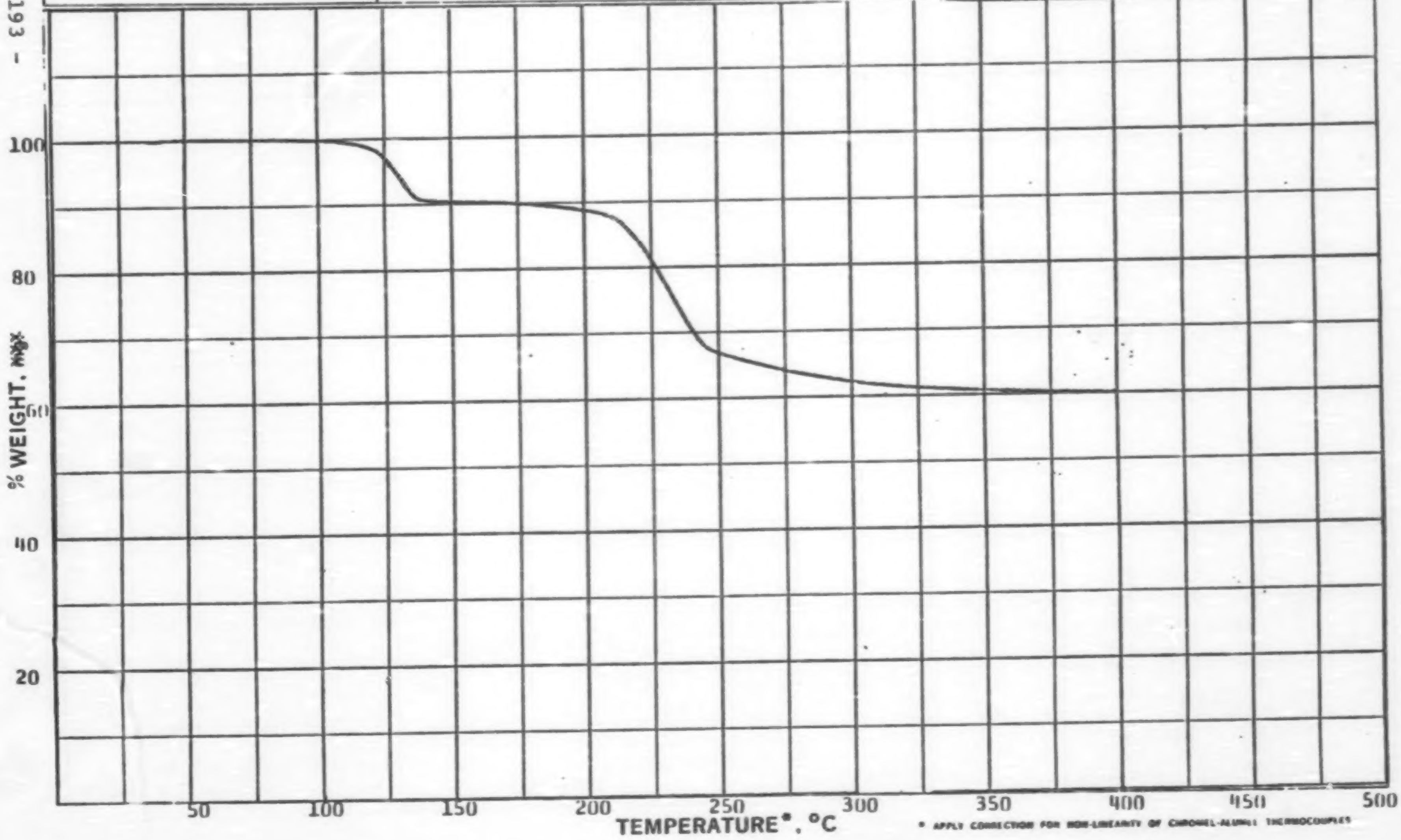
RUN NO. 1 DATE 11/27/74

OPERATOR RM

HEATING RATE 10 °C/min.

ATM. self-generated

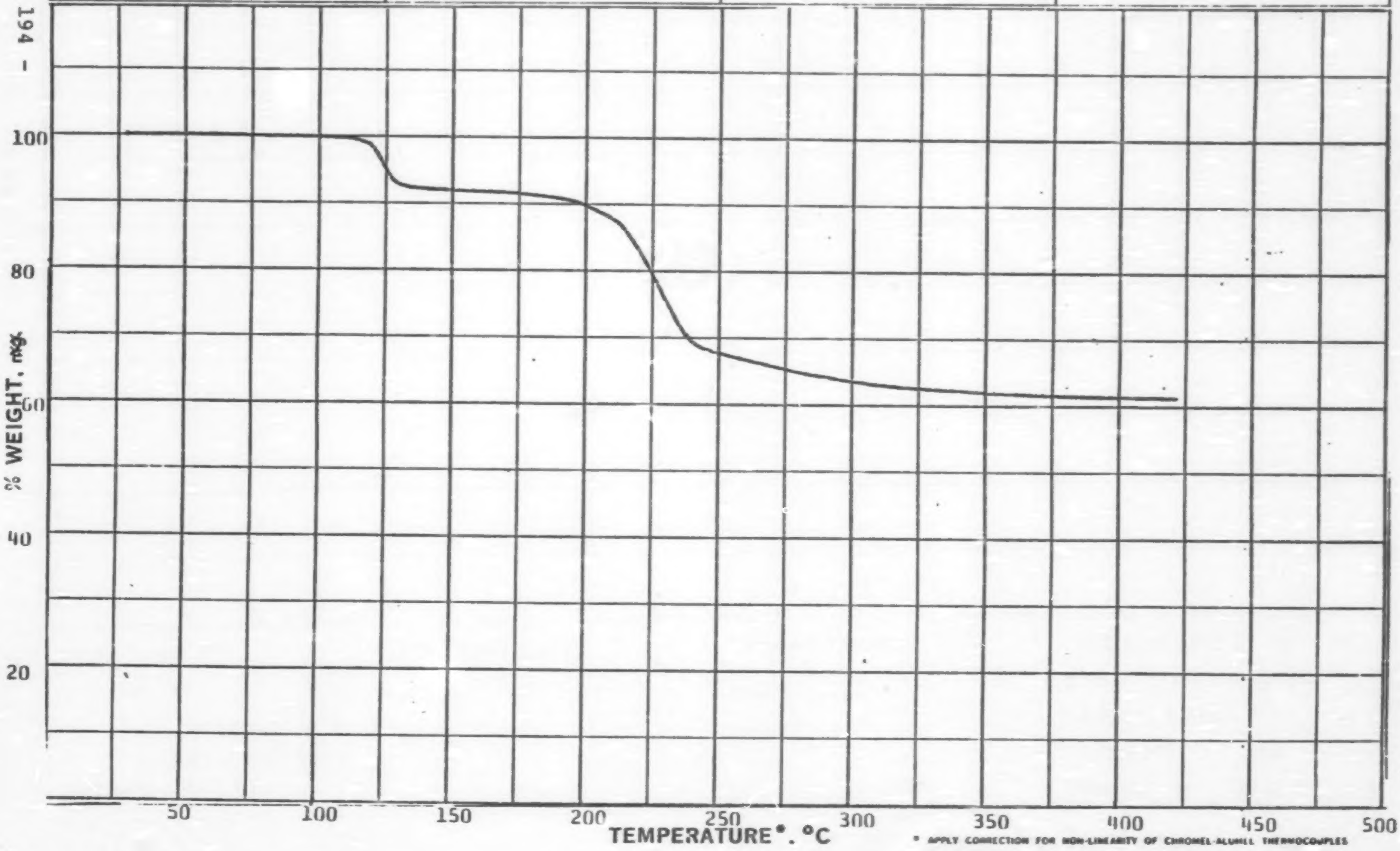
TIME CONSTANT 1 sec.



* APPLY CORRECTION FOR NON-LINEARITY OF CHROMEL-ALUMEL THERMOCOUPLES

DUPONT INSTRUMENTS

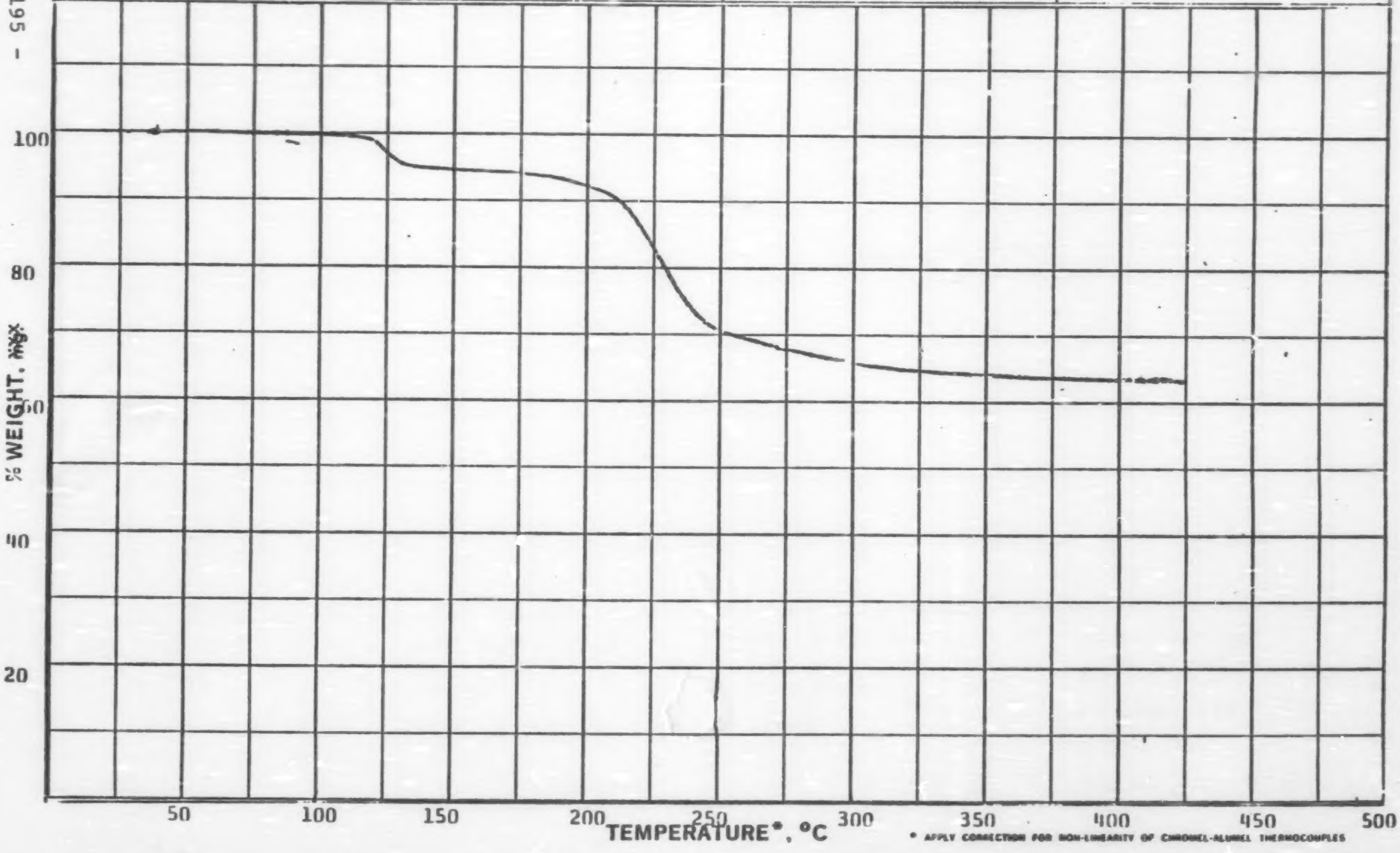
SAMPLE: Std #8 70.19% $MgSO_3.3H_2O$ 29.81% $MgSO_3.6H_2O$	**X-AXIS**	**Y-AXIS**	**RUN NO.** 1 **DATE** 11/25/71
	TEMP. SCALE 50 °C/inch	**SCALE** 2 mg./inch (SCALE SETTING X 2)	**OPERATOR** RM
Fig. No. 16	**SHIFT** 0 inch		**HEATING RATE** 10 °C/min.
			ATM. self-generated
SIZE 12 mg.	**TIME SCALE (ALT.)**	**SUPPRESSION** 50 mg.	**TIME CONSTANT** 1 sec.



* APPLY CORRECTION FOR NON-LINEARITY OF CHROMEL-ALUMEL THERMOCOUPLES

DUPONT INSTRUMENTS

SAMPLE: Std #9 79.34% $MgSO_3 \cdot 3H_2O$ 20.66% $MgSO_3 \cdot 6H_2O$	X-AXIS	Y-AXIS	RUN NO. 1 DATE 1'/27/7'
	TEMP. SCALE 50 °C/inch	SCALE 2 mg./inch (SCALE SETTING X 2)	OPERATOR RM
Fig. No. 17	SHIFT 0 inch		HEATING RATE 10 °C/min.
			ATM. self-generated
SIZE 14 mg.	TIME SCALE (ALT.)	SUPPRESSION 50 mg.	TIME CONSTANT 1 sec.



* APPLY CORRECTION FOR NON-LINEARITY OF CHROMEL-ALUMEL THERMOCOUPLES

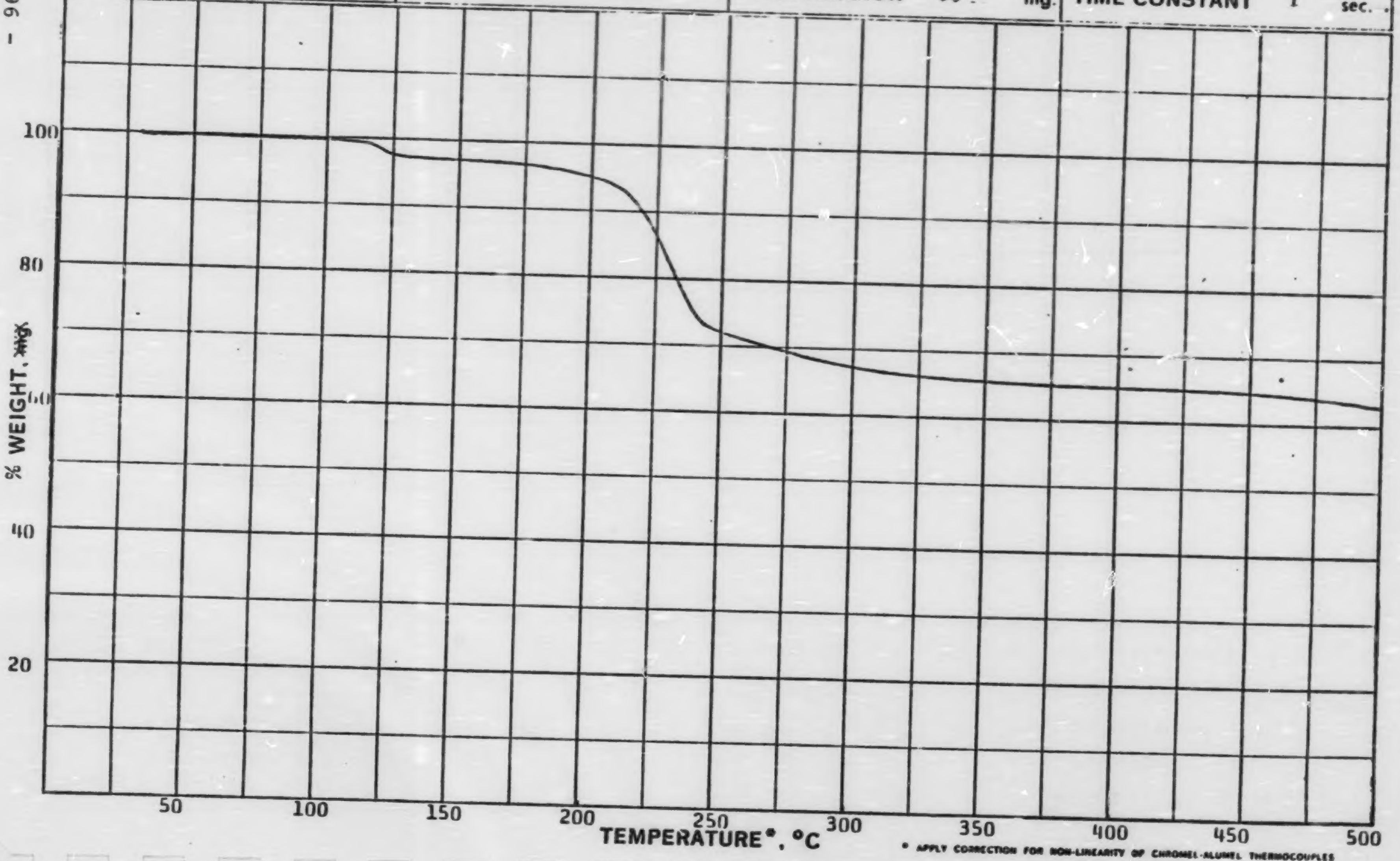
DU PONT INSTRUMENTS

SAMPLE: Std #10
90.18% $MgSO_3 \cdot 3H_2O$
9.82% $MgSO_3 \cdot 6H_2O$
Fig. No. 18
SIZE 16 mg.

X-AXIS
TEMP. SCALE 50 °C/inch
SHIFT 0 inch
TIME SCALE (ALT.)

Y-AXIS
SCALE 2 mg./inch
(SCALE SETTING X 2)
SUPPRESSION 50 mg.

RUN NO. 1
DATE 11/30/74
OPERATOR RM
HEATING RATE 10 °C/min.
ATM. self-generated
TIME CONSTANT 1 sec.



* APPLY CORRECTION FOR NON-LINEARITY OF CHROMEL-ALUMEL THERMOCOUPLES

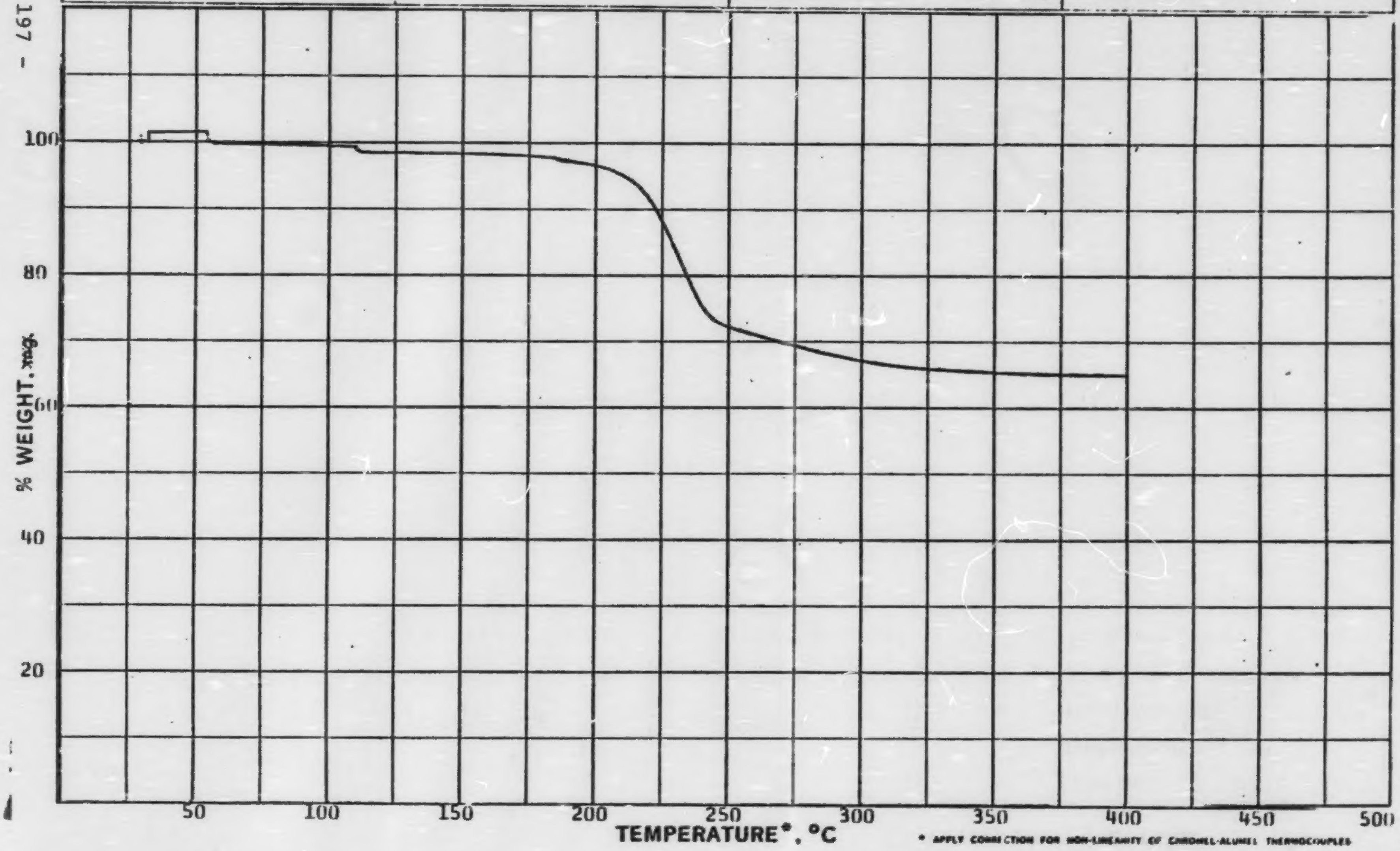
INSTRUMENTS

SAMPLE: Std #11
100% $MgSO_3 . 3H_2O$
Fig. No. 19
SIZE 14 mg.

X-AXIS
TEMP. SCALE 50 °C/inch
SHIFT 0 inch
TIME SCALE (ALT.)

Y-AXIS
SCALE 2 mg./inch
(SCALE SETTING X 2)
SUPPRESSION 50 mg.

RUN NO. 1 **DATE** 12/1/74
OPERATOR RM
HEATING RATE 10 °C/min.
ATM. self-generated
TIME CONSTANT 1 sec.



DUPONT INSTRUMENTS

SAMPLE: A mixture of:
44.3% $MgSO_3.6H_2O$
30.0% $MgSO_3.3H_2O$
25.7% glass beads
Fig. No. 20
SIZE 15 mg.

X-AXIS
TEMP. SCALE 50 °C/inch
SHIFT 0 inch
TIME SCALE (ALT.)

Y-AXIS
SCALE 2 mg./inch
(SCALE SETTING X 2)
SUPPRESSION 50 mg.

RUN NO. 1 **DATE** 12/1/74
OPERATOR RM
HEATING RATE 10 °C/min.
ATM. self-generated
TIME CONSTANT 1 sec.

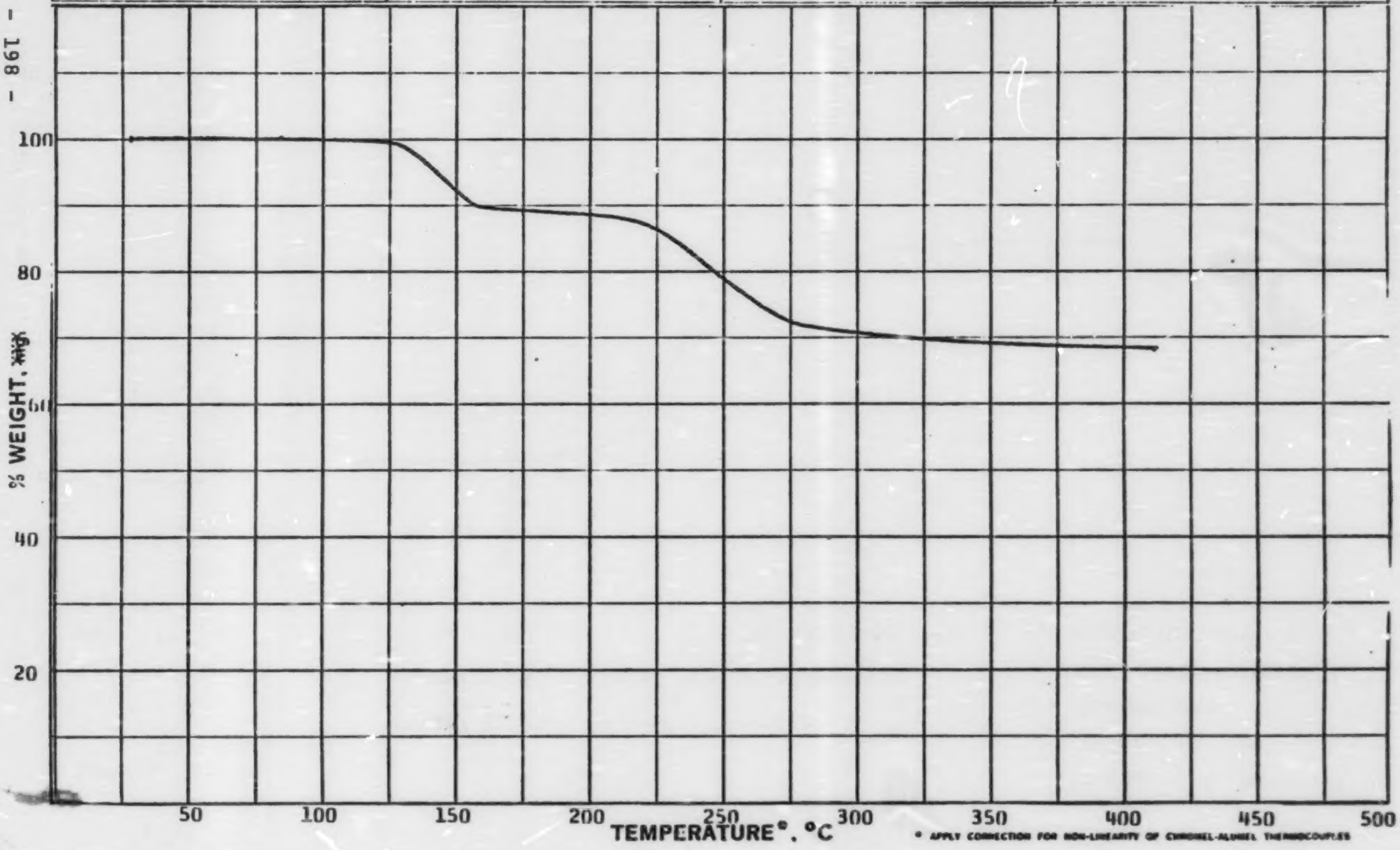


Fig. No. 21

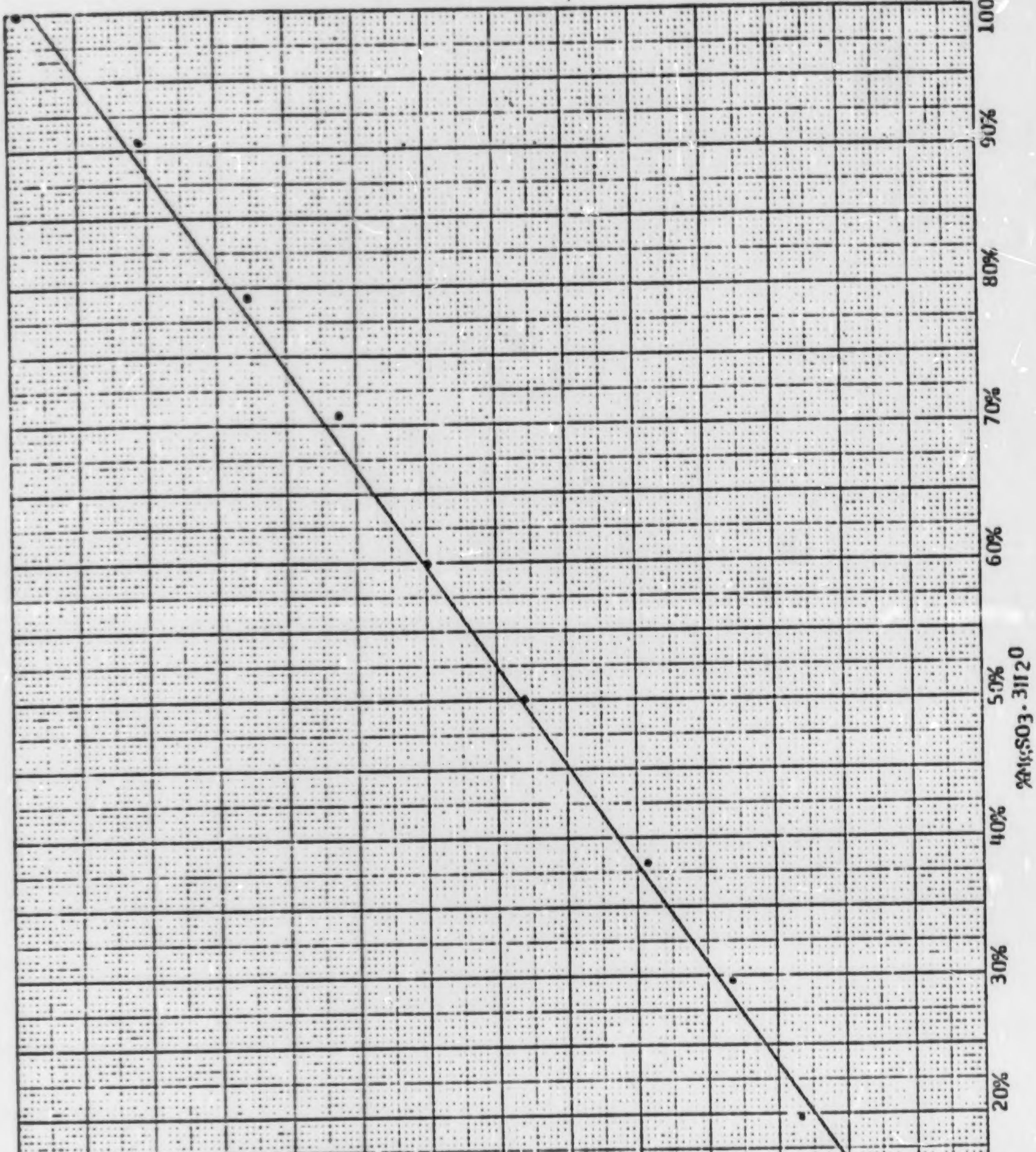
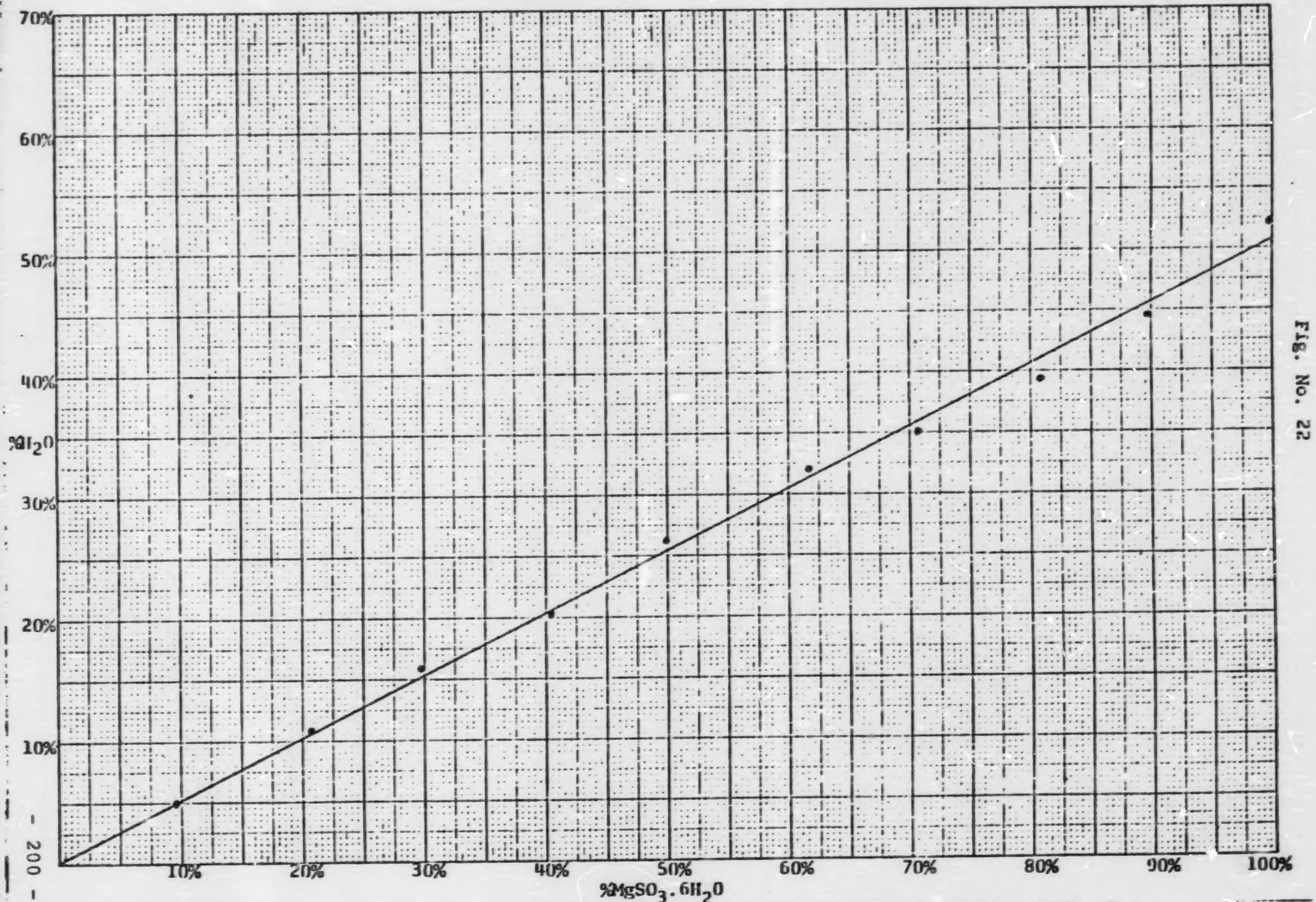


Fig. No. 22



APPENDIX 3

SULFUR OXIDE REMOVAL FROM POWER PLANT STACK GAS

Magnesia Scrubbing-Regeneration: Production of Sulfuric Acid

A Mass Spectrometric Study of the Thermal Degradation of $MgSO_3 \cdot 3H_2O$ and $MgSO_3 \cdot 6H_2O$

by

Prof. Leonard Dauerman
Department of Chemical Engineering and Chemistry
New Jersey Institute of Technology
Newark, New Jersey

Prepared for

Chemical Construction Co.
New York, New York

February, 1975

INTRODUCTION

In the Magnesia process for the removal of SO_2 from the flue gas and power plants, two hydrates of magnesium sulfite - the trihydrate and the hexahydrate - are formed. At a later stage these hydrates are calcined; MgO is formed which is recycled and, also, SO_2, which is converted into sulfuric acid.

It was the aim of this study to answer several questions pertaining to the calcining process: How do the hydrates decompose under vacuum conditions? Are trace amounts of H_2S formed in the thermal decomposition? How valid is the thermodynamic calculations on the partial pressure of SO_2 as a function and temperature. [1](K. Schwitzgebel and P. S. Lowell, "Thermodynamic basis for existing experimental data in $Mg-SO_2-O_2$ and $Ca-SO_2-O_2$ systems," Env. Sci. & Tech., 7, No. 13, 1147 (Dec. 1973).

The technique used in this study was mass spectrometry.

EXPERIMENTAL:

Samples of the hydrates were heated in the ionization source of the Finnigan Quadrupole 1015 mass spectrometer by utilizing the solid inlet probe. The temperature of the solid could be heated up to 400°C and the ambient pressure was approximately 10^{-6} mm. Samples were heated to a given temperature and spectra were taken after the spectral pattern had stabilized.

RESULTS AND DISCUSSION

Two series of runs are presented as representative of the data obtained. In Run #370, the spectra presented are those for the decomposition of $MgSO_3 \cdot 3H_2O$. Spectra are recorded from 35° to 400°C. The three parallel recordings on each spectrum represent data recorded at three different sensitivities

1:3:9. A similar set of records for $MgSO_3 \cdot 6H_2O$ are presented in Run #371 series of spectra. As a reference record, the spectrum of SO_2,H_2O is shown in Figure 1.

To determine whether or not there is a difference in the thermal decomposition between the two hydrates, the ratios of the peak heights of $\frac{SO_2}{H_2O}$ as a function of temperature were calculated.

In making these calculations data from an additional run for the trihydrate, and two additional runs for the hexahydrate were used. In Tables 1-5, the relevant data taken from the spectra are tabulated, and the value of the $\frac{SO_2}{H_2O}$ ratio is calculated. In Table 6, the ratios for the three runs for $MgSO_3 \cdot 6H_2O$ are averaged at each temperature, and the logs are calculated. In Table 7, similar operations are carried out on the two $MgSO_3 \cdot 3H_2O$ runs.

The log $(\frac{SO_2}{H_2O})$ as a function of temperature for the hydrates is plotted in Figure 1. The trend is similar. And in the light of the precision of the data, it can not be concluded that there is a difference in the way that the hydrates decompose.

Another question raised is whether H_2S, even in trace amounts, is observed. The peak characteristic of H_2S is observed at an m/e equal to 34. This m/e is observed in some spectra, for example, Run #370 at 400°C. The ratio of $\frac{m}{e}32/\frac{m}{e}64$ is 0.00547. In spectra Figure #1, for SO_2, where the intensity of m/e 64 is of a similar value the ratio is 0.00550. The fact that the difference in the ratio is the within the error in measuring the intensities means that the 34 is not an independent peak due to a separate species. It is, therefore, concluded that H_2S does not form even in trace amounts when the hydrates thermally decompose.

Note that SO_2 is observed at very low temperatures. In Run #370, it is observed at 35°C. But the pressure is ~ 10^{-6}mm. From Figure 2 in the previously cited paper by Schwitzgebel and Lowell, in which the equilibrium constant is plotted against temperature for the system $MgSO_3 \rightleftarrows SO_2 + MgO$, it is seen that at 35°C, the predicted value of the SO_2 pressure is 1.2×10^{-6} mm. This prediction is consistent with the observation made in this study, therefore, it is concluded that the equilibrium calculations are verified for the low temperature region.

CONCLUSIONS

1. Under vacuum conditions, $MgSO_3 \cdot 3H_2O$ and $MgSO_3 \cdot 6H_2O$ degrade similarly.
2. H_2S is not a thermal degradation product of the magnesium sulfite hydrates.
3. The thermodynamic calculations for the equilibrium $MgSO_3 \rightleftarrows MgO + SO_2$ were confirmed in the lower temperature region, 50-100°C.

MASS SPECTRA OF $MgSO_3.6H_2O$ & $MgSO_3.3H_2O$

RUN SERIES 370, 371, REFERENCE MATERIAL

The numerous mass spectra taken at the temperature intervals given in the table are not presented here since the pertinent data have been reduced in the subsequent tables and graphs.

Tables 1-5. CALCULATION OF $\frac{SO_2}{H_2O}$ INTENSITY RATIOS AS A FUNCTION OF TEMPERATURES FOR RUNS OF VARYING DETECTION SENSITIVITY FOR $MgSO_3 \cdot 3H_2O$ and $MgSO_3 \cdot 6H_2O$.

Table 1. $MgSO_3 \cdot 3H_2O$; Run #370; Detection Sensitivity, 1000 microamps.

Temp	m/e[1] 64	m/e 18	m/e 17	$\frac{SO_2}{H_2O} = \frac{64}{18}$
35°C	17	1500	480	0.01133
50	8	40	11	0.20
100	11	47	13	0.234
150	13	33	9	0.394
200	13	26	7	0.50
250	28	34	9	0.823
300	42	17	5	2.47
350	110	25	7	4.40
400	350	47	13	7.445

---

(1) m = atomic weight
e = charge

Table 2. $MgSO_3 \cdot 6H_2O$; Run #371; Detection Sensitivity, 1000 Microamps

Temp	m/e[(1)] 64	m/e 18	m/e 17	m/e 16	$\frac{SO_2}{H_2O} = \frac{64}{18}$
40°C	36	4610	1540	90	0.007810
50	250	15600	4900	320	0.01603
100	250	1830	580	40	0.1366
150	46	150	38		0.307
200	86	180	45		0.478
250	200	260	52		0.769
300	440	260	50		1.692
350	940	200	42		4.70
400	3250	350	68		9.286

---

(1) m = atomic weight
e = charge

Table 3. $MgSO_3 \cdot 6H_2O$; Run #372; Detection Sensitivity, 500 Microamps.

Temp	m/e[1] 64	m/e 18	m/e 17	$\frac{SO_2}{H_2O} = \frac{64}{18}$
35°C	5	450	150	0.0111
70	24	1674	558	0.01434
150	50	315	100	0.1587
200	33	80	22	0.4125
250	50	61	18	0.8197
300	98	63	18	1.556
350	207	35	10	5.914
400	423	72	18	5.875

---

[1] m = atomic weight
e = charge

Table 4. $MgSO_3 \cdot 3H_2O$; Run #373; Detection Sensitivity, 600 Microamps.

Temp	m/e (1) 64	m/e 18	m/e 17	$\frac{SO_2}{H_2O} = \frac{64}{18}$
35°C	- -	16	5	- -
50	13	500	130	0.026
100	10	350	80	0.0286
150	2	33	9	0.0606
200	2	32	8	0.0625
250	12	150	30	0.080
300	20	140	28	0.143
350	36	26	7	1.385
400	113	42	11	2.69

---

(1) m = atomic weight
e = charge

Table 5. $MgSO_3 \cdot 6H_2O$; Run #374; Detection Sensitivity, 750 Microamps.

Temp	m/e[1] 64	m/e 18	m/e 17	m/e 16	$\frac{SO_2}{H_2O} = \frac{64}{18}$
50	13	135	45		0.0963
100	74	4200	1400	91	0.1771
150	160	960	320		0.1667
200	55	160	36		0.3438
250	88	115	30		0.7652
300	200	98	22		2.564
350	520	65	17		8.00
400	850	100	27		8.50

---

[1] m = atomic weight
e = charge

Table 6. Ave. Ratio of $\frac{SO_2}{H_2O}$ as a function of T ($MgSO_3 \cdot 6H_2O$)

Temp (°C)	Run #371	Run #372	Run #374	Ave	Log Ave
35	--	0.0111	--	0.0111	-1.95
40	0.007810	--	--	0.007810	-2.107
50	0.01603	--	0.0963	0.0562	-1.250
70	--	0.01434	--	0.01434	-1.843
100	0.1366	--	0.1771	0.1569	-0.804
150	0.307	0.1587	0.1667	0.2108	-0.676
200	0.478	0.4125	0.3438	0.4114	-0.386
250	0.769	0.8197	0.7652	0.7846	-0.105
300	1.692	1.556	2.564	1.9373	0.287
350	4.700	5.914	8.00	6.205	0.793
400	9.286	5.875	8.50	7.887	0.897

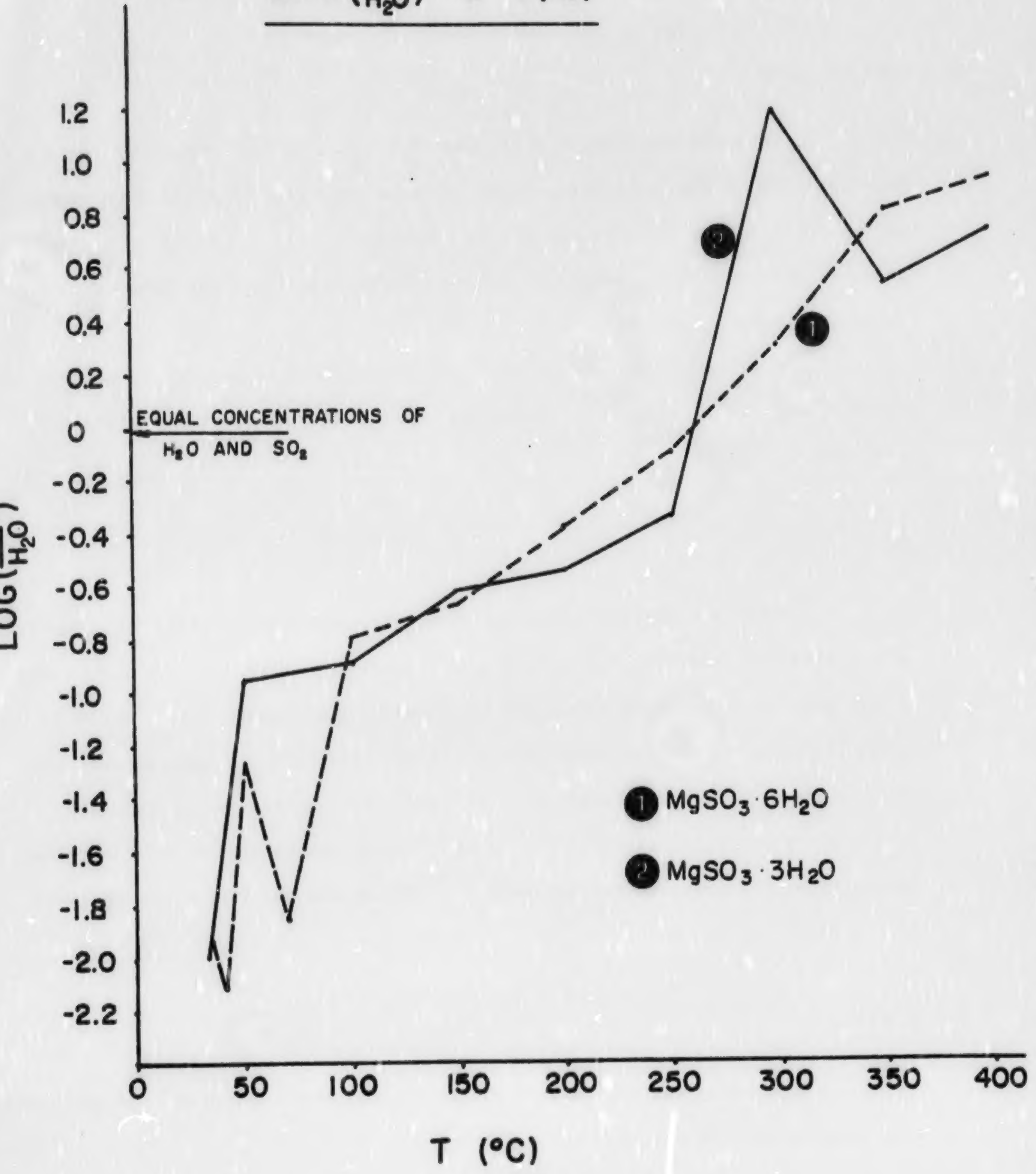
Table 7. Ave. Ration of $\frac{SO_3}{H_2O}$ as a function of T ($MgSO_3 \cdot 3H_2O$)

Temp(^{o}C)	Run #370	Run #373	Ave	Log Ave
35	0.01133	--	0.01133	-1.946
50	0.200	0.026	0.113	-0.947
100	0.234	0.0286	0.1313	-0.8817
150	0.394	0.0606	0.2273	-0.6434
200	0.500	0.0625	0.2813	-0.5509
250	0.823	0.080	0.4515	-0.3453
300	2.47	0.143	1.3065	1.161
350	4.40	1.385	2.893	0.4612
400	7.445	2.69	5.068	0.7048

Figure 2. Log $(\frac{SO_2}{H_2O})_{ave.}$ as a function of Temperature

for $MgSO_3 \cdot 3H_2O$ and $MgSO_3 \cdot 6H_2O$

$LOG\left(\frac{SO_2}{H_2O}\right)$ V. T(°C)



APPENDIX A. Simulation of Calciner in the Laboratory; The Testing of the Hypothesis that H_2S is Formed When the Burner Gas is Operated in the Fuel-Rich Region.

INTRODUCTION:

Since the previous study showed that H_2S is not a product of the thermal decomposition of the magnesium sulfite hydrates, and since H_2S had been reported as a product during calcination, it was conjectured that H_2S forms as a result of reactions between SO_2 and hydrocarbons present when the flame used to effect thermal degradation is operated in the fuel-rich region.

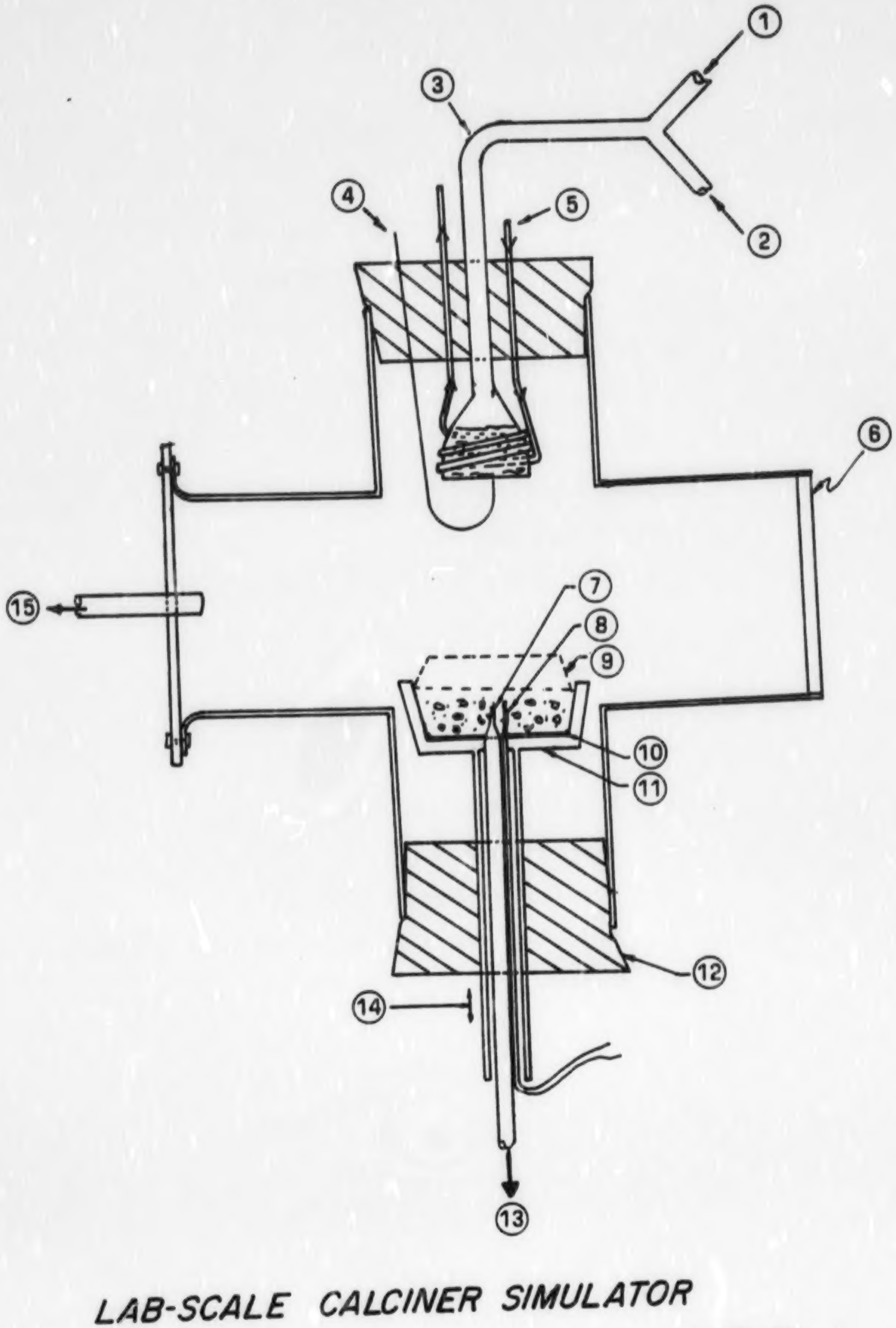
A laboratory set-up was designed which simulates many features of a calciner and, also, permits the immediate continuous analysis of volatilized products.

EXPERIMENTAL:

An inverted premix burner and a movable platform on which the sample is situated are housed within a pyrex glass pipe. See Figure A. In the center of the platform just above the sample is a quartz probe which leads to the Finnigan 1015 Quadrupole mass spectrometer. Thus, volatiles are continuously and rapidly characterized. In addition, the temperature in the region is sensed by a thermocouple. Also, the pressure in the "cross" can be varied and controlled. In practice, the pressure was maintained at 100 mm and the burner was operated in the fuel-rich region.

RESULTS:

This study was not carried out in a comprehensive manner. In the survey-type studies carried out, H_2S was not observed.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LAB-SCALE CALCINER SIMULATOR

APPENDIX A

LAB-SCALE CALCINER SIMULATOR

EXPLANATION OF NUMERALS

1. O_2 feed
2. Fuel feed
3. Premix burner
4. Lead for Tesla Coil ignition
5. Cooling water
6. NaCl window
7. Quartz probe
8. Thermocouple
9. Perforated stainless steel cover
10. Asbestos sheet
11. Stainless steel sample holder
12. Rubber stopper
13. To mass spectrometer
14. Position of thermal decomposition stage is adjustable
15. To vacuum

APPENDIX B. Effect of Wettness of Hydrates on Aggregate Formation During Calcining

INTRODUCTION:

The prevalent feeling had been that less aggregation, finer particles, resulted from the calcining of "dry" hydrates. On the other hand, it was thought in this laboratory that the volatilization of water during calcination should break up the particles.

EXPERIMENTAL:

Samples of $MgSO_3 \cdot 3H_2O$ were heated in air and vacuum to 500°C. Also, an extra-moist sample was heated in vacuum.

RESULTS:

Samples heated in air showed less aggregation than samples heated in vacuum. Extra-moistened samples heated in vacuum show less aggregation than the drier samples.

CONCLUSION:

Water decreases aggregation by explosive volatization, thus rapid removal by applying vacuum increases aggregation. Ideally, a wet sample should be rapidly heated to as high a temperature as possible.

TECHNICAL REPORT DATA

(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA-600/7-77-018	2.	3. RECIPIENT'S ACCESSION NO.
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7. AUTHOR(S) George Koehler		8. PERFORMING ORGANIZATION REPORT NO.
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15. SUPPLEMENTARY NOTES IERL-RTP project officer for this report is C.J. Chatlynne, Mail Drop 61, 919/549-8411 Ext 2915.

16. ABSTRACT The report gives results of a full-size demonstration of the magnesia wet-scrubbing system for flue gas desulfurization (FGD) on a coal-fired utility boiler. The system was designed to desulfurize half the flue gas from a 190-MW rated capacity generating unit firing 3.5% sulfur coal. The FGD installation was equipped with a first-stage wet scrubber for particle emissions control, followed by the magnesia unit. The FGD system was able to remove 90% of the inlet SO2 over 2800 hours of operation logged at the generating station. Its particle control capability was also demonstrated by reducing particle emissions to less than 0.01 gr/scf with the unit operated in series with an electrostatic precipitator. A test program, using only the wet-scrubbing unit for particle emissions control, achieved a collection efficiency of 99.6%. Magnesia was regenerated and recycled successfully. The SO2 produced during regeneration was used to manufacture commercial grade sulfuric acid which was marketed. Correlations were developed to determine SO2 removal for varying boiler loads and fuel sulfur content, and to control regeneration of acceptable alkali. Several other studies of the process technology and process chemistry were undertaken as part of the work.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group	
Air Pollution	Electric Power Plants	Air Pollution Control	13B	
Flue Gases		Stationary Sources	21B	10B
Scrubbers	Coal	Magnesia Scrubbing	07A	21D
Magnesium Oxides	Combustion	Particulates	07B	
Desulfurization	Particles		07D	
Sulfur Dioxide	Sulfuric Acid			

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